6e Symposium sur le traitement des eaux usées

16 & 17 novembre 1983

Hôtel Méridien

Montréal

6th Symposium on waste water treatment

November 16-17 1983

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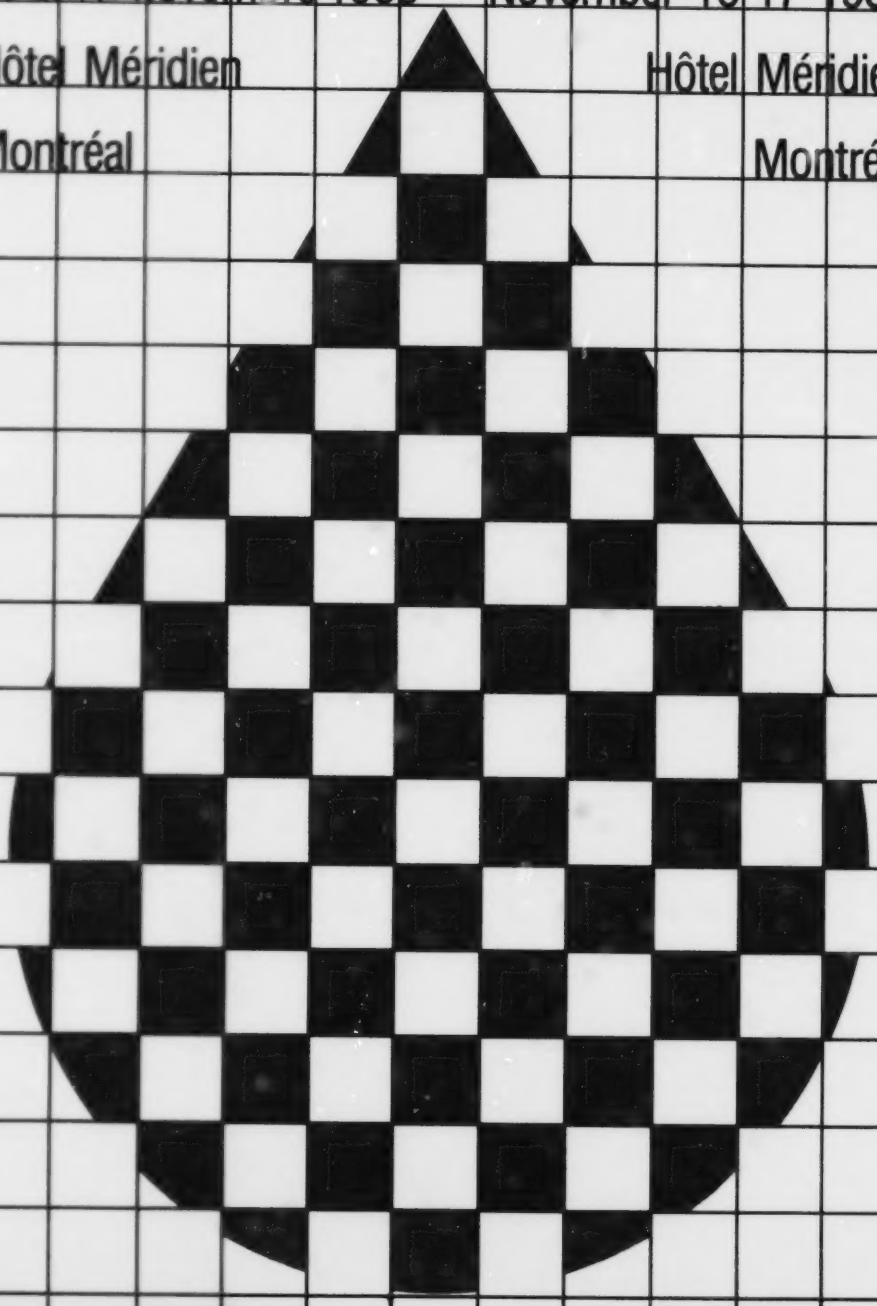

Comptes rendus

Proceedings

PROCEEDINGS

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Wastewater Treatment

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COMTES RENDUS

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des eaux usées

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Montréal

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PRÉFACE

Le comité d'organization et d'orientation du Symposium annuel est heureux de vous souhaiter la bienvenue à ce 6ième Symposium. Si cet événement annuel est devenu le premier forum d'échanges techniques dans le domaine de l'eau au Canada c'est grâce à votre support constant durant ces six dernières années. D'un éveil aux besoins nouvellement reconnus au Québec ce forum de transfert technologique est maintenant devenu un événement international. À cet égard nous avons reçu des offres de présentations de plus d'une dizaine de pays.

Ce Symposium s'adresse aux chercheurs, ingénieurs, opérateurs, technologues et, donc, spécialistes dans les domaines de la recherche, de la conception, de l'opération et de la gestion d'ouvrages de traitement des eaux usées. L'objectif premier du Symposium est de promouvoir et diffuser les connaissances et les techniques de traitement des eaux usées domestiques et/ou industrielles.

Le succès du Symposium depuis sa création et surtout le très haut niveau de l'échange technologique qu'il favorise ont permis à cet événement d'être reconnu sur le plan international comme un des grands forum scientifiques. Cette année particulièrement le symposium s'inscrit dans le cadre des activités de la "DÉCENNIE DE L'EAU", décrétée en 1980 par l'O.N.U.

Des spécialistes de plusieurs pays communiqueront le fruit de leurs expériences, de leurs connaissances et de leurs recherches dans le domaine du traitement des eaux usées.

Alain Jolicoeur

Ronald Zaloum

Lisette Provencher

Denis Lapointe

Claude Vergés

PREFACE

The organizing committee for this annual conference is pleased to welcome you to the 6th Symposium on Waste Water Treatment. If this annual event has become a leading forum in Canada for the exchange of technology relating to water, it is because of your continuing support during the last six years. A result of Québec's newly recognized needs, this forum has now become an international event. In confirmation, we have received offers of presentations from more than a dozen countries.

The Symposium is aimed at scientists, engineers, technicians, plant operators and specialists working in research, design and operation of waste water treatment facilities. The main objective of this symposium is to promote and share the know-how and techniques in the municipal or industrial waste water treatment fields.

Considering the success of the five previous symposiums and the high level of technology transfer, the Symposium has attained recognition as an international scientific forum. This year, the 6th Symposium subscribes to the "Decade of Water" designated by the UN in 1980.

Guest speakers from a variety of specialties and countries will share their most recent experiences and improvements in the waste water treatment field.

Alain Jolicoeur

Ronald Zaloum

Lisette Provencher

Denis Lapointe

Claude Vergés

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THE POLYSAR BIOLOGICAL OXIDATION TREATMENT PLANT

E.R. Fioroni
Polysar Ltd.

This paper deals with the Polysar Biological Oxidation Treatment Plant (Biox Plant), an industrial wastewater treatment facility located next to the St. Clair River at Polysar's Sarnia facilities.

Polysar produces synthetic rubber, lattices, and other petrochemicals at Sarnia. Many of the effluents from these production facilities are discharged to the Biox Plant for treatment. Some of these effluents have been pre-treated for the removal of oil and rubber which are reclaimed for other uses.

The design, construction and operation of this plant present many interesting and unique features about which to present a paper. This paper, however will only deal with three topics: 1) the close client/consultant working relationship during design and construction (which facilitates a 'fast track' approach); 2) the aeration system (which is the first of its kind in Canada); and, 3) the innovative training and commissioning program that was used for the Biox Plant.

The general layout, process design and some observations on the start-up and operation of the Biox Plant are also briefly discussed.

BACKGROUND

A Ministerial order was given to Polysar by the Ontario Ministry of the Environment to improve the quality of the waste stream that was being discharged to the St. Clair River from the Sarnia facilities. This was part of the overall program to improve the water quality in the Great Lakes.

In order to meet the quality requirements for discharge to the St. Clair River, it was decided that a central wastewater treatment plant would be constructed to treat most of the industrial waste streams.

The combined waste stream has a TOC (total organic carbon) concentration of 100 to 400 mg/l and a phenol concentration of 0.5 to 1.0 mg/l. The Ministry directive was to reduce TOC by 50% and phenol by 75%.

Pilot plant studies were carried out by Polysar for several years prior to design to determine the most suitable treatment process for their industrial waste. Biological treatment consisting of a conventional activated sludge system was found to be the most suitable means of treatment.

It was also confirmed during these studies that control of pH and the addition of nutrients would be necessary.

Flow rates and characteristics were found to be quite variable and dependent on production schedules. Start-up and maintenance operations in the industrial processes result in unpredictable waste flows. It was decided to size the plant for a nominal hydraulic capacity of 24 500 cu metres per day with a peak flow rate of 41 500 cu metres per day and a design strength of 10 000 kg TOC/day and 24 kg phenol/day.

DESIGN APPROACH

Marshall Macklin Monaghan Limited was awarded the design of the Biox Plant in July 1980. The target date for the start-up of the plant was July 1982, just two years later. This short period available for design and construction of the plant necessitated breaking up the project into packages that could be designed, tendered, and constructed individually and concurrently. This permited an early procurement of long-lead items and an advance start on site preparation and major excavation.

The Biox Plant project was divided into the packages which were prepared for tender call in the following order:

- preselection of aeration and clarification equipment
- aeration tanks and pre-treatment
- clarifiers
- electrical and instrumentation
- control building
- chemical feed and yard piping
- site works
- outfall works (marine and onshore)

Each of these packages was organized to provide minimal interference with other work on the project. The sequence of design was such that each package was tendered in time for construction to be completed by July 1982.

While the staging of procurement and construction is not unique on large projects, the speed with which it was achieved is noteworthy. Over 60% of the construction value had been tendered within 3 months of commencing design.

This 'fast track' approach necessitated good communications between Polysar and Marshall Macklin Monaghan. During the design of the major project packages, an office for the Polysar project team was established at Marshall Macklin Monaghan's Don Mills Office. This liaison office was staffed by various members of the project team from Polysar for three to four days per week on a regular basis over a 12 month period.

The liaison office enabled the client/consultant team to establish a close working relationship from the outset of the project, and ensured that the client was able to input actively into the design of the plant. Review of design concepts and detailed design was done mainly from the liaison office, and the resulting time savings in the review process ensured the rapid progression of the design.

It should be noted that the Polysar project team's purpose was not to take over the design of the plant, but to ensure the design progressed as quickly as possible, to minimize delays caused by client requested modifications to the design, and to ensure that each individual project package would fit into the overall project.

The key benefits that were realized from having the liaison office are:

- A close working relationship between the client and consultant teams.
- A faster design process, since review of concepts and designs was accomplished almost immediately.
- A design tailored to suit the needs of the client and the individuals who were to be responsible for operating the Biox Plant.
- The incorporation of engineering standards and specifications consistent with other systems at Polysar.

This close contact was maintained through the construction stages of the project as well. An engineer from the consultant's office was assigned to the project on a full-

time basis for the duration of construction. This provided a link between the design stage and the construction stage of the project. The liaison engineer attended weekly jobsite meetings and was responsible for liaison between the design and construction teams and co-ordination of shop drawing reviews and design modifications.

The use of the 'fast track' cooperative approach to design and construction proved to be very successful. The design began in July 1980, the first stage of construction began in April 1981, and micro-organisms were added to the aeration tanks to start-up the process in December 1982. The plant had been scheduled to start-up in September 1982, but a fourteen week strike by pipefitters delayed the start-up date until December.

The design and construction process took just two and one-half years. Had the entire plant been designed before any tenders were called the project could possibly have taken in excess of four years to complete. The best news is that the plant meets its objectives and was delivered at 10% under budget.

PROCESS DESCRIPTION

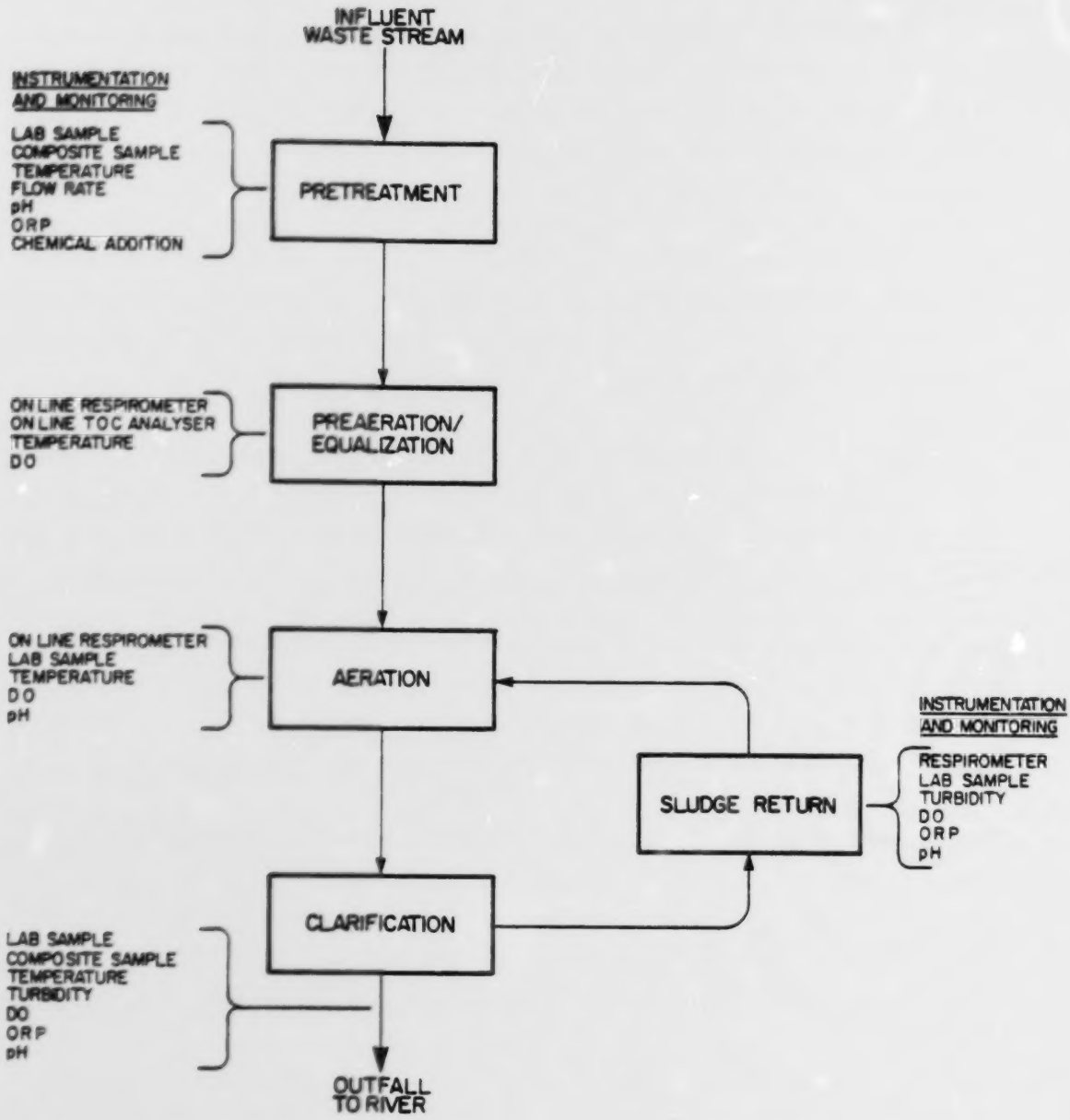
The major process components of the Biox Plant, as shown in Figure 1, consist of:

- o pre-treatment
- o pre-aeration/equalization
- o aeration
- o clarification
- o sludge return

The pre-treatment section consists of two pH control tanks and eight oil/solids separation tanks.

A mechanical turbine mixer in each pH tank ensures proper mixing of any chemicals that are added to the influent.

Either Sodium Hydroxide or Sulfuric Acid are used to control the pH levels of the influent. Rough adjustment of the pH is accomplished in the first pH tank, and final adjustments to pH are made in the second tank. pH monitors in each tank are used to automatically regulate the addition of acid or base by chemical metering pump control. Based on operating experience, the optimum pH for the biological process is 6.7. To achieve this, the pH in the pretreatment stage is adjusted to about 10.0.



MAJOR PROCESS COMPONENTS

FIGURE 1

If an influent enters the plant with a pH that is too high or low for proper adjustment by the pH control system, then an alarm signal is sent to the plant control centre. Appropriate action can be taken by the operator to keep the toxic influent from flowing to the activated sludge process.

Nutrients are also added to the plant influent at the pH tanks. Nitrogen and phosphorus, which are essential for efficient biological treatment are added as required to the influent. Monitoring the nutrients available in the influent ensures that only the required amounts of phosphoric acid (for phosphorus) and ammonium hydroxide (for nitrogen) are added.

Oil/solids separation is accomplished in eight separation tanks. Floating oil is collected in an oil scum trough and solids, which may accumulate at the bottom of the tank, are drained off by gravity. The oil/solids separation system is manually operated.

The agglomeration of oil present in the wastewater may be enhanced by the addition of ferric chloride, aluminum sulphate or polyelectrolytes, either individually or in combination. These chemicals can be added to the influent at the pH tanks. The rate of addition and combination used will depend upon the wastewater characteristics.

Pre-treatment chemicals are stored in a chemical tank farm area adjacent to the Control Building and are added to the system by chemical metering pumps. Because of the variations that can occur in the characteristics of the waste-water, only the addition of pH control chemicals is automatically controlled. Other chemicals are adjusted by the plant operators based on the results of daily samples taken from the waste stream.

An extensive on-line monitoring program, in conjunction with batch sampling keeps the operators informed of the characteristics of the influent from the time it enters the Biox Plant until it is discharged . Monitoring of the influent in the pre-treatment stage ensures that the chemicals added to the wastewater are kept to a minimum.

The pre-aearation/equalization tank follows the pre-treatment stage. This tank provides equalization of organic loads and flow and pre-aeration of the influent. The pre-aeration results in the partial oxidation of non-carbonaceous COD (chemical oxygen demand).

The normal liquid depth in the pre-aeration tank varies between 7.5 and 10 metres, and provides approximately three hours of hydraulic retention or about 3 200 m^3.

In the event that a toxic influent enters the plant, the pre-aeration tank can be used to contain this influent, and prevent it from reaching the aeration tanks, where it could damage the biomass. The contents of the tank can be discharged directly to the plant by-pass to protect the aeration tanks from undesirable influent.

The aeration tanks follow the pre-aeration tank in the wastewater flow through the Biox Plant. There are four aeration tanks, each one is 20 metres wide and 30 metres long with a 10 metre side water depth, giving a volume of 6 000 cubic metres per tank.

Aeration and mixing are accomplished by a jet aeration system. Each aeration tank has three distribution headers for the jet aeration system. Air is suplied to these headers by a central air blower. Each header has a recirculating pump which pumps liquid from the aeration tank through the header and back into the tank through a jet nozzle. These pumps are each rated at 236 l/sec (3 750 USGPM). Air is mixed with the liquid in this nozzle.

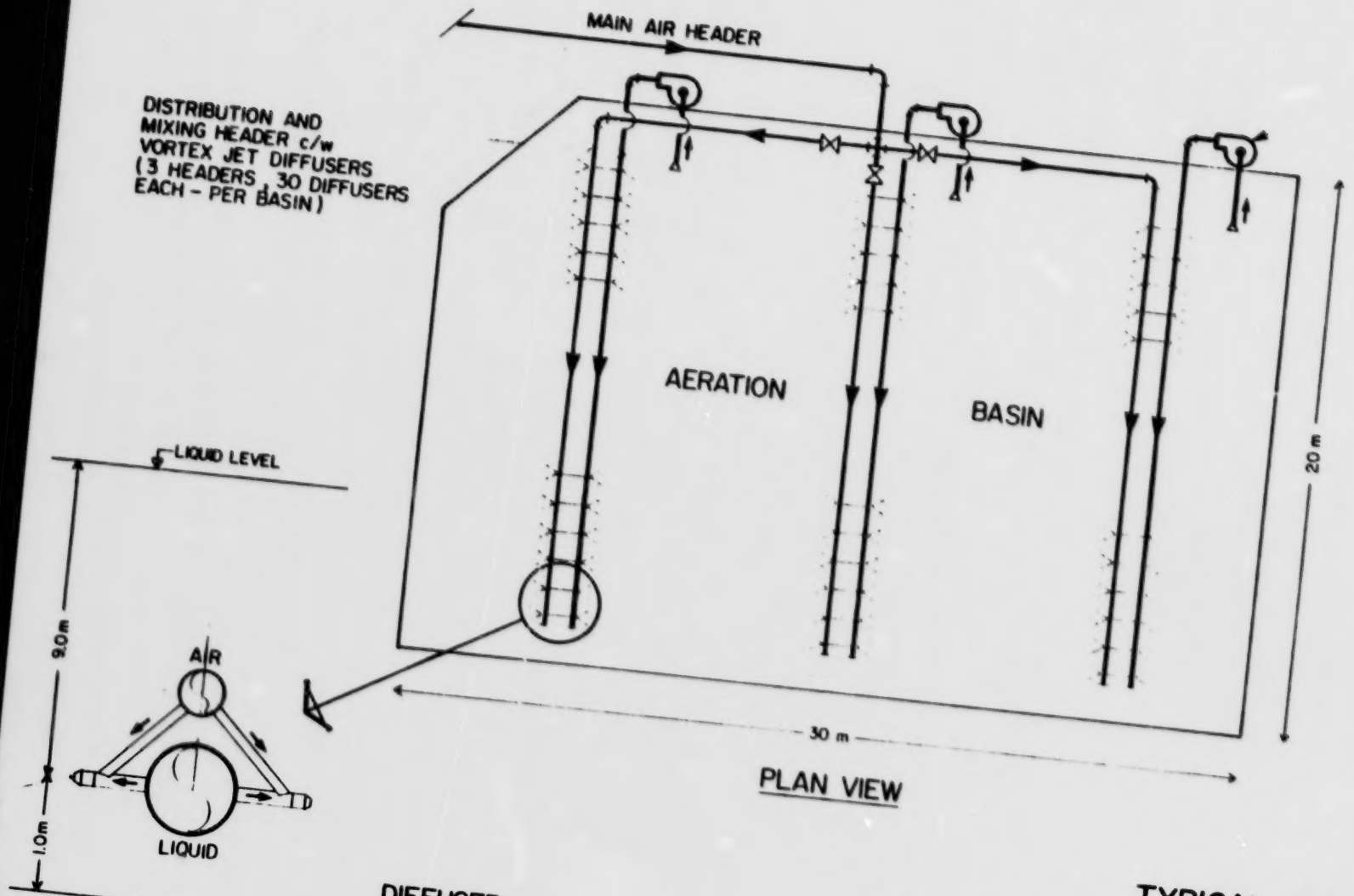
Oxygen transfer and mixing are provided by these jet nozzles. The discharge from the nozzle provides a roll-over type mixing in the tank. Oxygen transfer is obtained when the liquid and air are mixed in the nozzle, and by transfer from bubbles during their extended rise to the surface.

Each header has thirty nozzles on each side located approximately one metre above the floor of the aeration tank. The liquid depth in the tank is ten metres. The general layout of the jet aeration system is shown in Figure 2.

The four aeration tanks are arranged so that either complete mix or plug flow type regimes can be used.

With complete mix aeration, the influent is uniformly distributed to all four tanks. Each tank overflows a weir along one side of the tank. The complete mix system can handle surges in organic loading with minimal change in effluent quality. Return sludge from the clarifiers is equally distributed to all four tanks for complete mix.

Two types of plug flow are possible. With dual stage plug flow, influent enters one of two tanks, and flows into the adjacent tank through a submerged pipe. Four stage plug flow has influent entering only one tank and flowing, in series, through the other three tanks before being discharged from the aeration system.



TYPICAL AERATION BASIN
JET AERATION SYSTEM

FIGURE 2

Return sludge distribution from the clarifiers is controlled for plug flow, with most of the sludge being directed to the first tank, with a diminishing amount of sludge going to the subsequent tanks.

The plug flow configuration is superior to complete mix in its ability to avoid the short-circuiting of untreated wastewater during peak flow periods.

The aeration system is designed to maintain a DO (dissolved oxygen) level in the aeration tank between 2 and 4 mg/l. Air input to each aeration tank is based on the DO level in the tank. DO is monitored on a continuous basis. Air flow control is achieved by varying the blower capacity and throttling the airline to each tank. By throttling the airlines, air flow to each tank can be varied to suit plug flow conditions. Complete mixing can be achieved by liquid pumping alone, thus allowing the complete shut down of the compressors.

The pre-aeration tank utilizes a similar system for aeration.

The aeration tank effluent is discharged to four gravity clarifiers. The poor settling properties predicted by the pilot plant studies dictated a clarifier design with low hydraulic and solids loading rates. The design overflow rate for the clarifiers is 10 $m^3/m^2/d$.

Chemicals can be added to the clarifier influent to improve the settling characteristics. Aluminum trichloride, a waste product produced at Polysar, has been used in combination with 0.5 mg/l of anionic polymer to improve settling, and has avoided the need for other commercial chemicals.

The sludge blanket level in the clarifiers is controlled by variable speed recirculating pumps. Level sensors are used to control the speed and capacity of the pumps to maintain a stable sludge blanket level.

Three sludge pumps are provided to recycle settled solids from the clarifiers to the aeration tanks. Sludge pumping capacity can be varied from 50 to 200% of average design flow by means of motor speed control.

Process control of the plant is achieved by a data acquisition and control system. A central operators' control console located in the Control Building allows the operator to control and supervise the operation of the Biox Plant. To accomplish this, the control system will perform the following functions:

- Input monitoring of flow readings, level, temperature and chemical analyses;
- Signal conditioning and conversion to engineering units of inputs from monitoring equipment;
- continuous analogue and sequential digital control of chemical feed rates, aeration blower air flow and sludge recirculation rates;
- Alarm detection and annunciation;
- Display of operating data via an interactive display terminal;
- Long-term storage of process data.
- Hard copy print out of data in tabular and graphical form.

In addition to a very comprehensive range of automatic sampling devices, a continuously operating respirometer checks the treatability of the incoming waste.

AERATION SYSTEM

The selection of an aeration system is one of the more critical factors in designing an effective activated sludge system. An aeration system that does not satisfy oxygen transfer and mixing requirements will result in poor process performance. At the same time, aeration usually requires the highest energy input to the system, and any reduction in power requirements will translate directly into reduced operating costs for the plant.

The energy efficiency of the aeration system was, therefore, a prime consideration in the selection of an aeration system for the Biox Plant.

The size and shape of the property that the Biox Plant is built upon also played a key role in the design of the aeration system.

The combination of limited space, the need to provide for future expansion and the retention time that pilot plant studies had shown necessary for biological treatment made the use of large, shallow aeration tanks undesirable. The alternative chosen was deep basin type tanks, which would reduce the surface area required for aeration.

Analysis of the available aeration systems on the market suggested two types of aeration systems as being suitable for the deep tank configuration of the Biox Plant. These two types are the static mixer type and the jet mixer type.

The static mixer aeration system consists of air diffusers located throughout the bottom of the aeration tank. Both mixing and oxygen transfer are achieved by air bubbles emitted from the diffusers.

The jet mixer aeration system mixes air and wastewater together and discharges this mixture into the aeration tank through nozzles located on the bottom of the tank. These nozzles discharge horizontally. Oxygen transfer occurs both when the air is mixed with the wastewater in the nozzle and when the bubbles from the jet nozzle rise through the wastewater. The air/wastewater discharging from the nozzle and the rising action of the air bubbles provide the mixing action in the aeration tank.

Tenders were received for two static type systems and two jet type systems. The tenders were analyzed and compared on the basis of capital, operating and maintenance costs.

The system that was found to be most suitable for the Polysar Biox Plant was the jet type aeration system manufactured by Aerocleve-Clevepak.

This system, while more expensive to purchase than the static type systems, provided significant savings in operating costs, and was judged to have the lowest maintenance requirements. The equivalent annual cost for the system over a twenty year life proved to be the lowest of all the systems.

The jet aeration system also provides a built-in back-up capability, since if a recirculating pump fails, then air supply can provide mixing in the tanks, and vice-versa.

The jet aeration system is well suited to a deep tank configuration. It provides an oxygen transfer efficiency as high as 3.6 kg O_2/kW hour (6 lbs. O_2/hp hour), and the jet action ensures sufficient mixing in the aeration tank.

The high oxygen transfer efficiency means that less air is required for this type of system than for the static mixer type system. The static mixer had almost twice the air requirements of jet aeration.

The jet aeration/deep tank system for the Polysar Biox Plant is the first installation of its kind in Canada.

The jet aeration system requires a high volume, low pressure air supply (24 500 m^3/h at 102 kPa). Due to the variable quality and quantity of the wastewater to be treated, it was also desirable to have an air supply that could be varied to suit the air requirements in the aeration tanks.

Tenders were received from six different manufacturers for an air supply system to suit the selected aeration system. An analysis and comparison similar to the one performed for the aeration system was done for the air supply system.

The basic type of air blower systems evaluated were vertically-split centrifugal, horizontally-split centrifigual, and variable diffuser vane centrifugal. Volume control on the vertically and horizontally-split centrifugal blowers is by means of suction throttling. The variable vane blower controls discharge volume by varying the angle of the vanes on the impeller.

The variable diffuser vane blowers were found to be the most economical when compared on the basis of equivalent annual capital and operating costs over a 20 year period. The blower chosen for the Biox Plant was a Cord Turbo variable diffuser vane blower. The efficiency of this type of blower remains relatively constant throughout the entire operating range, at approximately 80%. The capacity can be regulated between 45% and 100% with a constant delivery head by changing the pitch of the variable diffuser vanes.

Three of these blowers are used. One blower can provide low volume air requirements, and two blowers in parallel provide the peak volume requirements. The third blower is standby.

This is the first installation of this type of blower in North America, although these blowers have been successfully used for wastewater treatment in many European countries.

The three blowers are housed in a separate building next to the aeration tanks. No other equipment is located in this building due to the noise levels that are generated when the blowers are in operation. The acoustical design of the blower building ensures that the noise levels outside the building are minimized.

Reliability and maintenance were carefully evaluted for both the blower and aeration systems. Although there were no similar installations in Canada, the installations in other parts of the world were evaluated and through comparisons with conditions in Sarnia, an estimate of reliability and maintenance was made.

To-date, both systems have been operating as they were designed and no major problems have occurred. The jet aeration/variable diffuser vane blower combination is an economical and reliable method of aerating a deep tank activated sludge system.

While this type of system will not be suitable for all aplications, it is certainly worth considering as an option during the preliminary design stages of an activated sludge system.

The 'fast track'/cooperative approach to design and construction has proven itself to be a useful approach for the Polysar Biox Plant. This approach was successful in having the design and construction complete in two and one-half years, however, the project did not end there. The operator training and commissioning program was the next stage in ensuring that the Biox Plant was put into operation in a short time, at minimum expense and with minimum start up problems.

TRAINING PHILOSOPHY

The capability to train operators exists. The questions that were difficult to answer at the start were: Can we "educate" people to take advantage of the training opportunities available but most of all, can we teach upper Management what an operator is? And if we are able to teach them, will they believe us?

It is well known that strange ideas prevail in the outside world that wastewater operators are not required to have broad technical knowledge.

Having realized that we were dealing with a powerful union, our first obstacle was to convince the Union that we requird experienced and skilled operators in the wastewater field. Our second objective was to communicate this information in a favourable manner to upper Management. At the beginning, it was very difficult for them to understand the validity of our convictions. They could not visualize "waste" requiring individuals that needed extensive experience and adequate training.

Looking back, the task of educating people not involved in the process itself was the most difficult hurdle to overcome. How did we achieve this goal? We gave some very sad examples of how major bioxidation plants were unable to meet the desired requirements because of lack of proper pre and post start-up training.

After long and difficult communication meetings, the authorization was given by upper Management and the Union to come up with an extensive and satisfactory training program.

At this point, we divided our training requirements into three parts:

- o Pre-Start-up Training
- o Start-up Training
- o Post Start-up Training

PRE-START-UP TRAINING REQUIREMENTS

It was decided that the Biox Plant would require six operators, six operator assistants, two maintenance people, a Process Engineer, a Foreman, a Secretary and a Plant Manager.

The following was the timetable suggested:

o	Plant Manager	on-site	2 years in advance
o	Processs Engineer	on-site	18 months in advance
o	Foreman	on-site	15 months in advance
o	Operators	on-site	12 months in advance

1. We initially recommended to hire all the six operators from outside Polysar which could give us a solid baseline of experience and knowledge. Due to union pressure, we were only able to obtain three outside operators. These three individuals all possessed University Degrees, 6-20 years of wastewater experience and were very well versed on activated sludge problems and start-up procedures.

2. Our next step was to hire the remainder of the required operators and assistants within Polysar. We decided to request the top individuals that were interested in working in an Activated Sludge Plant. The interest was very high as over 100 applied for the job. That made it much easier in obtaining top operating staff within the plant.

3. One of the problems associated with wastewater treatment plants is that operating personnel do not stay on the job for very long, causing discontinuity and lack of knowledge derived from minimal on-site experience. We had to develop a sound, progressive training program to ensure that our operators would consider their job a permanent one.

4. The decision was made that the only way the operators could leave the Biox Plant was if they personally choose to post out. This decision enabled us to ensure the permanency of "Biox Operator" and allowed us to design a solid progressive certification program.

5. In Polysar, laboratory technicians have a different job description than process operators, and one cannot perform the other's duties. Therefore, we had to design a job description that would fit both categories and at the same time be satisfactorily accepted by the Union. We did this by convincing Management and Union that the operator will be efficient, if he personally performs the required tests and does the actual interpretation. This dual responsibility will make his job more enjoyable.

6. Since process control will be achieved by a data acquisition and control system (DACS) which is to include a central operator's console and process interface cathode ray tube (CRT) units, the operators had to be properly and thoroughly trained to operate and understand the computer system.

   A training program consisting of three full days of theory, and three full days of practical application was given four months prior to the targeted start-up date. This program would help the operator in:

   (a) understanding the console operation and how to properly interpret data through process control.

   (b) minimize the sense of mystique and terror that the monstrous but simple machine gave to individuals that were not properly trained.

Every operator, assistant operator, foreman, engineer, secretary, all instrument people and maintenance personnel took the course.

Although the course was an excellent tool, we strongly recommend that this type of training be carried out during or just before actual start-up. Our course was repeated three weeks before start-up.

7. At this stage, having permission to hire three outside operators and to obtain the top candidates from within the plant, we established a permanent position for our personnel and were successful in outlining the responsibilities of the operators within their job description.

One key item was still missing: we had to convince Management that we required these individuals at least six months to one year before the plant actually was operational. During this time frame, we would ensure proper operation and maintenance of the wastewater system, in accordance with the design to achieve maximum plant performance and efficiencies of treatment.

We established certain requirements necessary for a professional and effective training program. The trainer would be able to:

(a) correctly define and categorize problems and determine whether they can be solved through training;

(b) analyze the tasks performed at the plant in order to determine the training needs;

(c) determine the type and level of knowledge required for each task;

(d) specify instructional objectives which will be taught to the employees;

(e) develop evaluation activities to determine the progress of students and the effectiveness of the training itself;

(f) choose the best method of instruction for specific objectives;

(g) choose and create effective media for different types or levels of objectives;

(h) communicate with other people on the job site;

(i) deliver a lesson effectively;

(j) revise and evaluate training materials in order to remain current in a dynamic profession;

(k) develop a training schedule.

8. The first six months would be dedicated to achieving the above. The next six months would require training the remaining operating personnel and a training schedule was prepared.

There was one major concern: nine out of the twelve operators had no biox experience. We decided to install a major pilot plant so that our operators could use it as a training and learning tool. Two identical separate systems were built.

Each system consisted of:

One	-	400 litre feed tank (with agitator)
One	-	78 litre aeration tank, the aeration system was under continuous agitation with air flow control
One	-	21 litre clarifier with rotating mechanicsms, 1-5 rpm
Two	-	parastaltic pumps (0-500 ml/min.) for feed and return sludge

With the aid of the Pilot Plant, the operators would learn how to define a problem and find its possible solution. They gained expertise in how to operate instruments and laboratory machinery, but most of all they were able to understand that an activated sludge plant will offer many problems that will require interpretation gained by a solid technological approach (i.e. actual microscopic analysis and interpretation). The operators very quickly learned NOT TO OVERREACT.

9. Our last pre-start-up objective was to send the operators to an operating activated sludge plant, within our area, for a few days to give them some practical experience. Every operator was sent to an activated sludge plant for three days with very positive results. It was at this time that we realized how effective our training up to that point had been. Our operators had more basic knowledge than most of the other operators they talked to. This was primarily because we were able to obtain our personnel one year before start-up but also because we had made our job description a permanent one. From this, we made a serious commitment to ensure our operator's continuity and job security within the Biox.

Summarizing, let us analyze the pre-start-up achievements:

(a) defining the basic need for manpower;

(b) obtaining personnel with practical experience and vast wastewater knowledge;

(c) instituting a security in the job requirement to ensure the operators will not move;

(d) ensuring that the job description included a clause for operators to have dual responsibilities: laboratory and process;

(e) developing training schedule;

(f) obtain all operating personnel six months to one year before actual start-up;

(g) construct and operate a miniature pilot plant;

(h) visit other Bioxidation Plants within specified area;

(i) training.

START-UP TRAINING REQUIREMENTS

The start-up procedure, at this point, was well established through the use of the Pilot Plant. The operators, with the aid of the Pilot Plant, were able to actually experiment and collect data in determining which method would be most advantageous.

After analyzing all the various facets of our needs, we decided that our operators should be involved in overseeing the final phase of the construction. This would satisfy two major requirements:

(a) it would give the operators hands-on-experience of all equipment (actually see the internals), lines, laboratory facilities and all pumps. By comparing in detail the drawings with the actual construction, the operators would be able to point out outstanding errors and possibly have an early input in design changes. In short, they will be able to learn how the plant physically operates.

(b) It would ensure that all design and construction parameters were met satisfactorily before the plant became direct responsibility of Biox team.

The following is how we assigned responsibilities during this period:

Operator 1	-	Clarifier area
Operator 2	-	Underground and overhead piping and pipe gallery
Operator 3	-	Laboratory and building
Operator 4	-	Mechanical areas and chemical area
Operator 5	-	Aeration and preaeration
Operator 6	-	Pretreatment, oil separation and instrument room

We also had to train other Polysar personnel to understand what a Biox Unit was and its purpose within the overall plant structure.

We initiated sound and professional short presentations outlining the plant design and trying to explain what the consequences could be if effluent conditions were not met. Our objective was to instill in our other units not only a sense of appreciation for the high technology required and a devotion needed within Biox but also to ensure a sense of respect and loyalty to our plant. We had to radically change ideas and conceptions that are vividly exemplified by the common statement: "Don't worry, send it to the Biox Plant, they will take care of it."

These presentations last 30 minutes followed by one hour discussions repeated four to five times during the first year of operation.

POST START-UP TRAINING FORMAT

To minimize labour management relations, it was mandatory to outline a certification program and make it part of the job requirement. Certification ensures the following:

1. It will allow operators to maximize their potential.

2. It will ensure that the operators will perform with the most innovative and modern technology.

3. It will motivate the operators to continuously learn new technology, operation strategy and will also enable them to more clearly understand the intricate problems associated with the Biox Plant.

4. It will ensure an overall "good feeling" amongst the entire crew arising from self confidence and ability.

5. It will save Polysar a tremendous amount of dollars:

    i) dollars derived froma better D.O. control (less electrical costs);
    ii) dollars derived from an overall improved control (less maintenance costs);
    iii) dollars derived from ability to interpret results and make necessary changes that will avoid major problems;
    iv) dollars derived from an overall better environment and increased work production.

The certification of operators will be based on experience, education and examination with provision for limited substitution of education and experience with each other under an approved system.

Formal education can be supplemented by approved training courses, seminars, etc. for which C.E.U.'s (continuing education units) are allocated.

Considering the fact that in Ontario, there is no mandatory or voluntary certification program as of yet, it is noteworthy that we, at Polysar, have developed a four phase training program that will eventually lead to a complete certification when it is established in this province.

Let us look at the four phase training program developed:

Phase I - Basic

This phase is intended to give the staff a review of Grade 12 and first year college subjects that they will require in the every day operation of the plant. The subjects are:

Mathematics	(16 units)
Chemistry	(8 units)
Biology	(2 units)
Physics	(2 units)
Microbiology	(3 units)

Each operator will be writing a pre-test on each subject. The result of the test will enable us to place them in the appropriate level. Some operators will not require this first phase and will automatically go to Phase II. This will be all in-plant training. In order to institute an effective in-plant training program, existing technical personnel were utilized in development and delivery. We expect an operator to complete Phase I within three to six months.

Phase II

During this phase, they will be taking a basic laboratory course, developed and implemented within the plant.

They will also start having hands-on-experience through the digestion and understanding of the Operation and Maintenance Manuals.

The objective of this phase is to introduce them to the laboratory technology (theory and practice) and also allow them to slowly digest the operation of the plant via a practical approach. This should take approximately six months to one year.

Phase III

During this phase, the operators will be concentrating on an advanced laboratory course where they will learn to accurately interpret results and make necessary changes. During this phase, they will attend the "APPLICATION OF ON-LINE RESPIROMETRY AND OTHER ANALYTICAL INSTRUMENTATION PROCESS CONTROL COURSE."

We feel that this course is imperative since our operational strategy is based on respirometry and D.O. control. At this course, they will be exposed to a practial approach in understanding the science of respirometry and settling properties.

They will then attend a 60-hour wastewater course at our Community College. This course is a prerequisite for Level IV. This phase should take approximately one year to eighteen months.

Phase IV

At the completion of Phase III, the operator is now ready to undertake the Ministry of the Environment Courses offered in Toronto. We have chosen eight courses to be taken over a period of three years.

MINISTRY OF ENVIRONMENT COURSES

	Course	Experience	Education	CEU	Pre-Requisites
1.	Basic Sewage Treatment Operation	6 Months	Lambton College	3.0	Lambton College
2.	Activated Sludge Workshop	1 Year	Lambton College	3.0	Course No. 1 + Lambton College
3.	Primary Treatment & Digestion Workshop	1 Year	Lambton College	3.0	Course No. 1 + Lambton College
4.	Basic Water Treatment Operation	1 Year	Lambton College	3.0	Lambton College
5.	Surface Water Treatment	18 Months	Lambton College	3.0	Course No. 1 + Lambton
6.	Pump Operation Workshop	1 Year	Lambton College	3.0	Lambton College

MINISTRY OF ENVIRONMENT COURSES

	Course	Experience	Education	CEU	Pre-Requisites
7.	Laboratory Skills	2 Years	Lambton College	3.0	Course No. 1 or 3 and Course No. 2 + Lambton College
8.	Advanced Water/ Wastewater Treatment	3 Years	Lambton College	3.0	Course No. 2 or Course No. 5 + Lambton College

At this stage, the operator will possess the experience supplemented by adequate training to be considered a "qualified wastewater operator." This will be to his advantage when certification becomes compulsory.

Summary of the four phase training program:

Phase I

Basic Knowledge - Mathemetics (16 units)
- Chemistry (8 units)
- Biology (2 units)
- Physics (2 units)
- Microbiology (3 units)

UP TO 6 MONTHS

Phase II

- Introductory laboratory course
- O&M exposure with hands-on-experience

6 MONTHS TO 1 YEAR

Phase III

- Advance laboratory course
- Attendance of course: Application of On-Line Respirometry and Other Analytical Intrumentation to Process Control
- Wastewater Course at Community College (60 hours)

1 YEAR TO 18 MONTHS

Phase IV

- 8 Ministry Courses, duration 1 week each

2-3 YEARS

PLANT OPERATION

The plant started up on December 8, 1983 and immediately (after 10 days), the desired M.O.E. requirements were met. The T.O.C. was reduced by 50% and the Phenolics by 80%.

The results in the last six months are outstanding, even with major problems, our operators were able to quickly implement corrective actions to prevent any major disaster. Our maintenance costs are minimized, our energy costs are optimized and we did not require any outside chemical addition.

We can conclude at this time that the training program coupled with the flexible and well designed plant were the two key ingredients for a successful, operational plant.

CONCLUSION

Effective training is not an accident but is a systematic approach. The mechanics of planning the training itself leads to both a pre-start-up program and continuing training for company personnel. We feel that proper training is the single most important tool in a well operated unit.

Constructing, expanding and upgrading wastewater treatment facilities present numerous employee related problems as well as technical difficulties. Prior planning and training will pay off in the reduction of problems that will occur during the initial start-up period and the subsequent operating period. To ensure a complete cost effective and productive operation, the following four factors should be considered:

1. **PROPER DESIGN** to allow for process flexibility and controllability.

2. **CLOSE CLIENT/CONSULTANT WORKING RELATIONSHIP** during design and construction which facilitates an efficient and timely design.

3. **INSTRUMENTATION** (on-line and laboratory) to allow for process information to be readily available.

4. **SOLID AND ESTABLISHED TRAINING PROGRAM** so that the operator can learn to properly interpret process information and apply it to a flexibly designed plant.

ANAEROBIC PRETREATMENT FACILITY ALSO PROVIDES SLUDGE DISPOSAL CAPABILITY AND SOURCE OF RENEWABLE ENERGY FOR FOOD PROCESSOR

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In 1980, a multi-national food processing company decided to locate a medium-sized potato processing plant in France. Part of the overall project included waste treatment facilities which had to be ready upon completion of factory construction in 1981. The treatment facilities were designed in 1980, construction also commenced in 1980, and was completed in mid 1981, together with the factory.

The waste treatment facilities are comprised of raw waste pumping station, vibrating screens, potato waste hoppers, ADI-BVF,* aeration nitrification complete with clarifier and sludge return, control building and ancillaries including biogas handling equipment and automatic samplers. Phase 1 of the system has been operated since August, 1981; Phase 2 facilities will be added at such time as the factory proceeds to implement Phase 2 production capacity. Figure 1 provides a schematic flow diagram for the Phase 1 facilities.

Design loadings are shown in Table 1 as well as the assumed design removals to meet the final effluent values imposed by the French authorities. An interesting feature was the imposition of nitrogen limit.

Key design criteria for the unit operations are supplied in Table 2.

Additional details of the various operations and processes are supplied below.

Unit Operations and Processes

Raw Wastewater Pumping Station.

Two, submersible, torque flow pumps rated at 80 m^3/h each and capable of handling 15 cm solids deliver the wastewater to the screens. The pumps are wired for duty/standby arrangement with both pumps coming on to cope with peak flows. Each pump discharge is independently piped to a vibrating screen.

*Trademark of ADI Limited

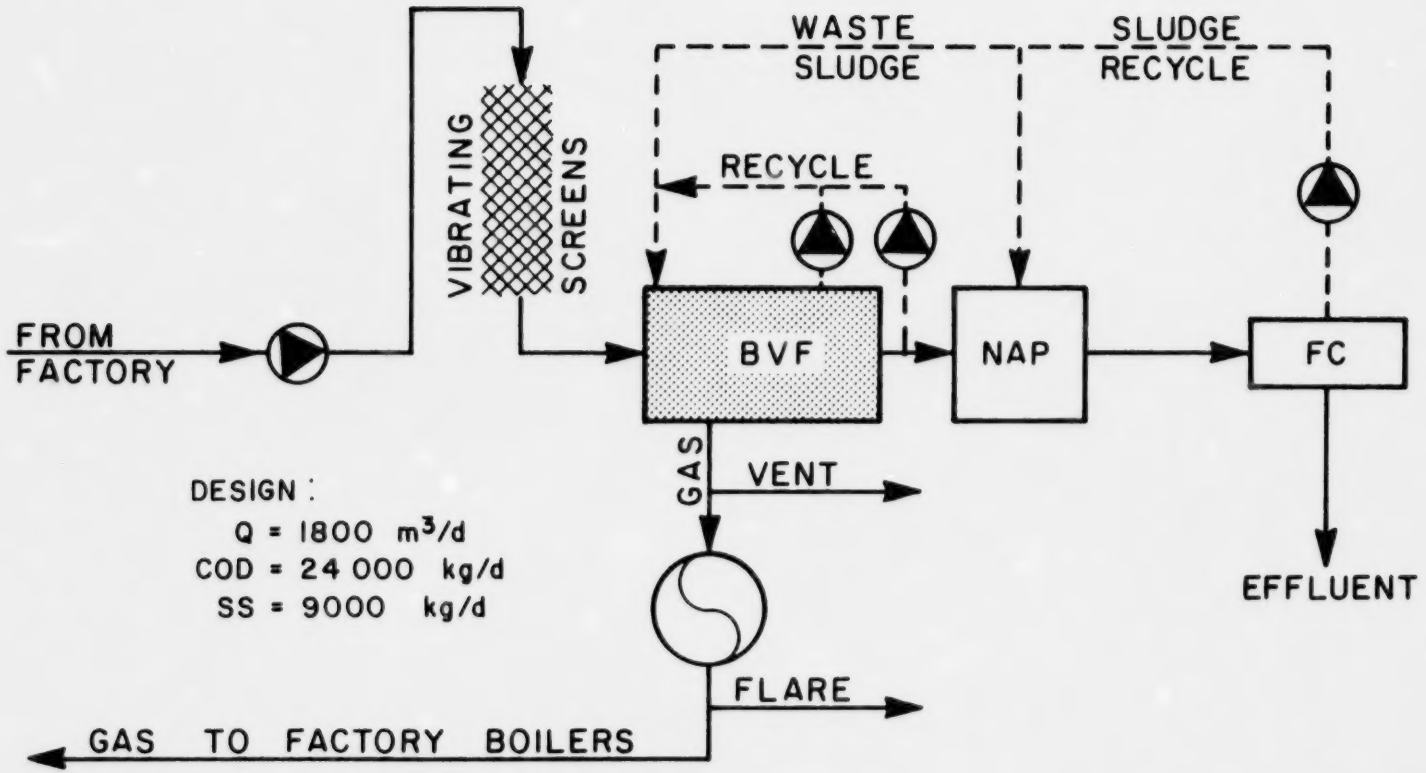

FIG 1: SCHEMATIC OF TREATMENT FACILITY

TABLE 1 - DESIGN PLANT LOADINGS AND REMOVAL

Location	Parameter	Load kg/d	(ppm)	Removal, %	
Plant	Q	1800 m^3/d			
BVF (screened effluent)	COD	24 000	(13 300)	90	
	NOD	3 430		-	
	TKN	750		-	
	SS	9 000	(5 000)	90	
	VSS	7 650		-	
Aeration-Nitrification	COD	2 400	(1 330)	50	
	BOD	1 200	(667)	83	
	TKN	750	(416)	90	(conversion to nitrate)
	Fat	100	(55)		
Final Effluent Limits	BOD	200	(111)		
	COD	1 200	(667)	95	(overall)
	TKN	75	(42)	90	(overall)
	SS	200	(111)		
	pH			5.5-9.5	

TABLE 2 - DESIGN CRITERIA

ADI-BVF

COD loading, average	1 $kg/m^3.d$
Detention time, average	13 d
SS loading, average	0.4 $kg/m^3.d$
Temperature, minimum	15-20°C
Time lapse to first sludge removal	4 years

Aeration Nitrification

F/M (combined oxidation-nitrification)	0.25 or less
DO minimum	2 ppm
SRT, minimum (at 10°C)	15 d
Recycle ratio	0.75-1.5
Volumetric loading for TKN	0.21 $kg/m^3.d$
Oxygen required, kg/d	1.5 x BOD + 4.57xTKN
Sludge production	0.3 kg/kg BOD_r

Final Clarifier

Hydraulic loading rate, average flow	12 $m^3/m^2.d$
Side water depth	4 m

Vibrating Screens.

There are two, rectangular, 10 mesh vibrating screens each rated at 4500 m^3/d. The screens are electrically-linked to the supply pump motors. The screens are mounted atop 100 ton hoppers to collect solids and peel waste. The screened effluent flows by gravity from the screens to the ADI-BVF.

Chemical Addition.

A self-contained lime silo-feeder is in place. It can apply a dosage of up to 500 ppm although the average assumed dosage was 200 ppm.

ADI-BVF.

This anaerobic fermenter has dimensions of 57 m x 109 m x 6 m deep providing a volume of 26 000 m^3 at maximum water level. It is basically of earthen construction although interlocking sheet steel piles were used to make the sides vertical for 3 m. The steel piles served two functions: to penetrate an impervious stratum lying just beneath the floor level, and to facilitate construction of a concrete cap which served as a neat wall to attach the cover. The bottom 3 m of depth of the ADI-BVF was constructed of earth with 3:1 side slopes.

A floating flexible membrane cover together with a central, floating gas collection pipe is used to collect the biogas. The gas recovery blowers apply a negative pressure to the floating cover which causes it to lie flat on the liquid surface. The cover is furnished with two access hatches and several sampling ports which are also used to monitor sludge depth. A flexible connection on the gas pipe allows the liquid level to routinely vary by up to 1-2 m.

There is an inlet manifold header to introduce the feed and an exit header to remove the effluent. A system of headers, pipes and pumps is provided to recirculate or waste sludge; sludge can be removed while the fermenter remains fully operational. There is also provision to recycle the effluent by means of two pumps each rated at 150 m^3/h.

The blower building contains two 5.5 kW electric-motor-driven blowers, each with a capacity of 400 m^3/h. They can apply 0- 50 mm wg suction pressure to the cover and 0.3 bars pressure discharge. Biogas is delivered to the process boilers, fitted with a natural gas and a biogas burner ring, where it is burned to produce process steam for the factory. The blower building also houses safety equipment, oxygen sensors, gas meter, emergency flare and vent and ancillary equipment.

A photograph of the ADI-BVF appears in Plate 1.

Aeration Nitrification Basin.

This membrane-lined basin has an operating volume of 3600 m^3. It is an earthen pond having dimensions of 38 x 38 m at the surface, 2.5:1 side slopes and a liquid depth of 5 m.

Aeration is provided by either of two, 185 kW self-aspirating jet aerators. The submersible motors help to maintain temperature in the basin in winter to foster nitrification. A DO sensor is used to pace the aerators both for purposes of process control and energy conservation.

Final Clarifier.

A suction clarifier was used in order to return sludge in a fresh state. Its dimensions are 18 m x 9 m x 4 m SWD.

PLATE 1: ADI-BVF IN OPERATION

Sludge is recycled to the aeration basin by either of two centrifugal pumps; one is rated at 150 m^3/h while the other is variable from 30-200 m^3/h capacity. Interconnecting piping permits activated sludge to be wasted into the ADI-BVF.

Control Building.

The main control building contains the pumps associated with fermenter and activated sludge, a laboratory and an office. It also has an electrical room and storage space.

Sampling and Instrumentation.

Three automatic samplers are provided - one for screened effluent, one for anaerobic effluent and one (refrigerated) for the final effluent. Composite samples are collected over a 24 h period. The samplers are activated by flow meters. The screened effluent composite is also used to supply a fermenter pilot; this pilot is 1/100 000 the size of the ADI- BVF, and it is proportionately batch-fed daily with wastewater collected on the previous day.

Probes and readout equipment are provided for fermenter influent and effluent pH and temperature, and for DO in the aeration basin.

Raw wastewater flow is measured by a Palmer-Bowlus flume while final effluent is measured by a 150 mm Parshall flume.

Plant Start-Up

During the construction of the treatment system, operating manuals were prepared, complete with startup procedures.

The specified startup procedures for the anaerobic portion were as follows.

1) Fill ADI-BVF with 25°C water.
2) Add 2000 m^3 of municipal digester sludge initially, plus 100 m^3/week thereafter.
3) Commence the gradual addition of factory effluent or fine- ground raw potatoes, for the 3-4 month startup period.
4) Monitor TA (total alkalinity), VA (volatile acids), pH, biogas and COD, and if stress is evident or imminent, increase alkalinity and reduce or maintain the feed input at a constant level until results permit an increase in feed.

Seeding the ADI-BVF with 2400 m^3 of municipal digester sludge at 4% solids was completed in July, 1981, shortly before the plant startup. However, filling the BVF with 25°C water was interrupted because of a construction problem. This took several months to resolve during which the scheduled startup procedure was not followed. The program to import 100 m^3/week of seed sludge to facilitate startup was not implemented. Consequently, by November, 1981, the volatile acids were slowly rising, and pH and COD removal were decreasing. This indicated that the system was under stress. However, by year end the construction problem had been resolved and it was then decided to return to the planned startup procedure which was called the second startup.

Steam was added to the factory effluent to increase the temperature of the fermenter effluent from 17°C up to 25°C. Chemical addition of lime and sodium bicarbonate was implemented to boost alkalinity and the addition of

seed sludge from a municipal digester was implemented; (1000 m^3 in December, 500 in January, 300 in February and 200 in March, all at 4% solids). The parameters monitored on the ADI-BVF indicated that with these modifications the fermenter was no longer under stress and the COD removal had increased to 70%.

The second startup being almost complete, the heating and the addition of seed sludge were stopped and the addition of lime greatly reduced by March, 1982. The performance of the fermenter continued to increase, yielding 86% COD removal even though the influent load fluctuated greatly (from 6500 to 17 000 kg/d of COD) while its temperature fell to 21°C. By May, 1982, the addition of lime had been discontinued, and the fermenter was performing nicely. Due mainly to the fact that sodium bicarbonate was left over from startup, 200, 100, 100, 100 and 50 kg/d were added in August, September, October, November and December, respectively; no chemicals have been added in 1983.

The factory was shut down for the complete month of July, 1982, for employee vacations, plant repairs and major washup. During this period there were no chemicals added, neither was there recycling nor sludge mixing in the ADI-BVF. Recycling was started up one week before plant startup and the BVF liquid level was lowered in order to start progressively feeding the aeration system in advance of factory startup to preclude bulking. The fermenter was not affected by this shutdown period as evidenced by the COD removal which continued to improve upon resumption of factory operations in August, 1982.

Operation

The results for the year ending April, 1983, have been very good for the entire waste treatment facilities, especially for the ADI-BVF anaerobic fermentation system. The entire facilities were operated by 1-2 operators on one shift, 5 d/week (corresponding to factory production of two shifts for 5-6 d/week). The authorities are pleased concerning effluent quality. The owner is also pleased over the fact that the monetary value of the biogas energy more than offsets the operating and maintenance costs of the entire treatment facility - from the factory wall to the receiving water.

The authorities require the reporting of a large number of tests. Consequently, there was a considerable amount of data available to assess plant loadings and performance. Each point shown on the graphs is a monthly average value of tests conducted by the factory.

Figure 2 presents the temperature, detention time and loading on the ADI-BVF. The influent was generally around 30°C whereas the effluent ranged from 25°C in warmest weather to 20°C in winter. The effluent temperature was below 20°C for several days, the lowest recorded daily value being 17.5°C. However, the pilot plant was constructed of steel (uninsulated) and located in an unheated room, consequently its temperature fell to a minimum daily value of 13°C, yet, it continued to perform well. During coldest weather last winter the ambient temperature fell to -5°C at night, with 2°C during the day.

Commencing in October, 1982, the detention time had fallen to 13.65 days, (based on processing days only) and it continued to fall thereafter; reaching its lowest value, 11.55 days, in March, 1983.

The COD loading was around 0.6 kg/m^3.d until January when it began to

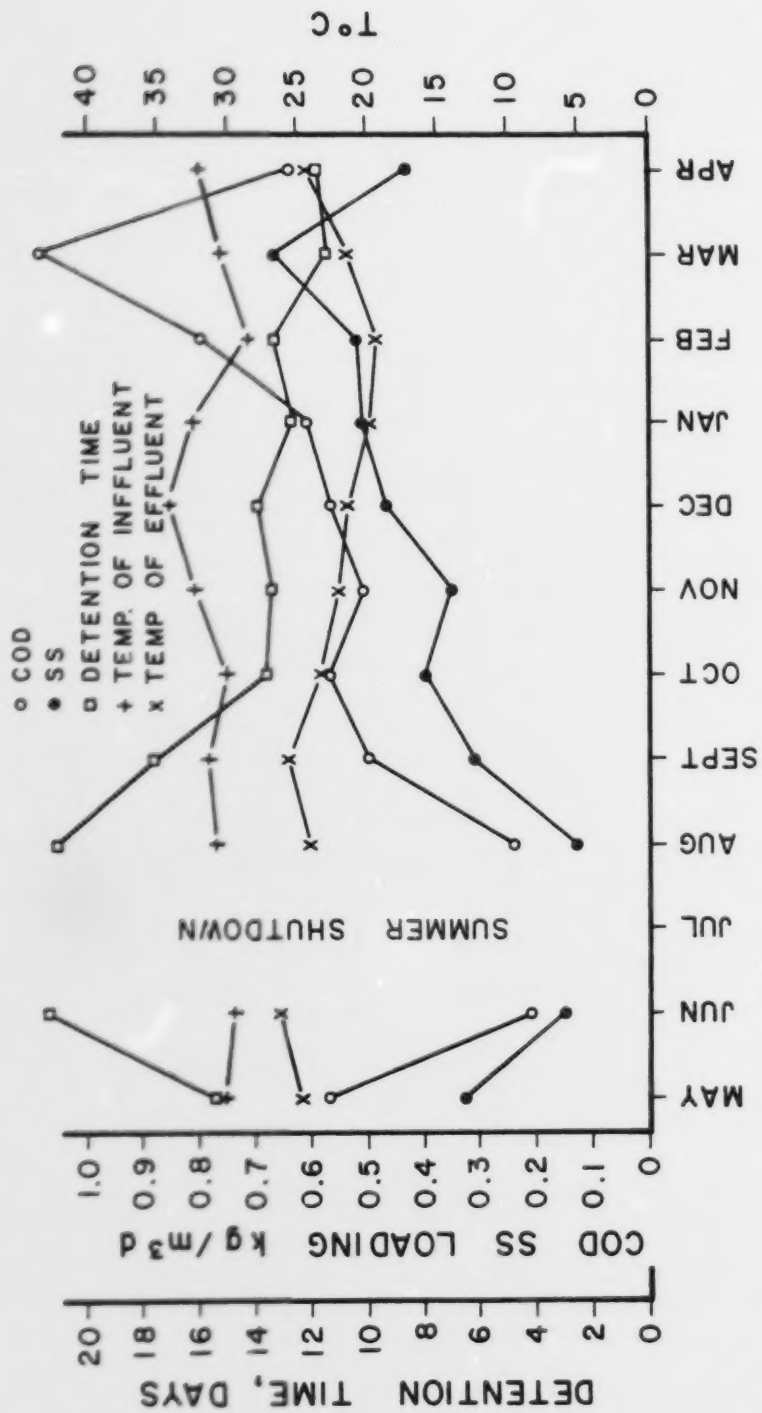

FIG 2: TEMPERATURE, DETENTION TIME AND LOADING ON ADI - BVF

rise sharply, reaching a maximum value of 1.09 in March before falling off.

The SS loading gradually rose after fall startup, reaching a maximum value of 0.67 kg/m^3.d in March, 1983. It should be noted that the COD and SS loadings (and removals) include the load applied as a result of wasting activated sludge into the BVF. However, the waste sludge is rather insignificant, amounting to approximately 3% of the COD load and 5% of the SS load. However, since percentage removals shown in Figure 3 were calculated on a concentration basis, rather than mass basis, this resulted in essentially no change in the case of SS, and a slight reduction (less than one percentage point) in the case of COD, since the COD of the waste activated sludge was less than the COD of the factory influent, expressed in ppm.

Figure 3 presents COD, BOD and SS removals in the anaerobic fermenter along with methane production. After initial startup the COD removal had reached 85% in May, 1982; it continued to rise after fall startup in spite of increasing loading, decreasing temperature and decreasing detention time, settling in at approximately 91%. The same pattern is true for BOD removal except it settled at a somewhat higher value, 95%.

Suspended solids removal, too, displayed a similar pattern, reaching a typical removal of 93%.

Methane production rose and fell monthly, generally according to the COD load entering the fermenter. A calculation showed that, when the weekend production of biogas was taken into account, the recovered biogas amounted to approximately 0.32 m^3/kg COD removed, at STP.

Turning to the aerobic portion of the plant, Figure 4 presents the loadings, retention and temperature in the aeration basin. As the performance of the fermenter improved in the spring of 1982, the BOD loading fell from 0.5 to 0.1 kg/m^3.d where it remained till April. The detention time normally ranged between 2.0-2.5 days. This is greater than the 2.0 day design value, in spite of higher-than-design flows, as a result of using the fermenter as equalization storage, (such that flows entering aeration on weekends were approximately the same as on week days).

The ammonia-N loading held fairly steady at 0.05 kg/m^3.d for the whole year. The F/M (VSS basis) was also rather constant throughout at approximately 0.02. The MLSS was allowed to vary between 3 000-4 500 ppm.

The SVI ranged from as high as 207 in June to as low as 53 in December. The large difference is not just due to the variation in settleability of the sludge; the higher values generally correlate with the higher MLSS concentrations due to the nature of the SVI calculation.

Aeration pond temperatures reached 20°C in summer months and fell to a low of 10°C in January.

The performance of the activated sludge portion of the plant was also considered good, judging by the BOD and ammonia removal as may be seen in Figure 5. The SS removal was typically around 80%, except for lower values in August, September and October, following the summer shutdown.

The plant influent and final effluent concentrations for BOD, COD and SS are supplied in Figure 6. Typical monthly averages were 7-8 000 ppm for influent COD although June was only 4 300 ppm, due to low factory production, while March was 10 100 ppm due to extra waste losses in plant. Influent BOD following summer shutdown typically averaged 4-5 000 ppm with three months exceeding 6 000 ppm. Final effluent BOD's were typically in the 10-15 ppm

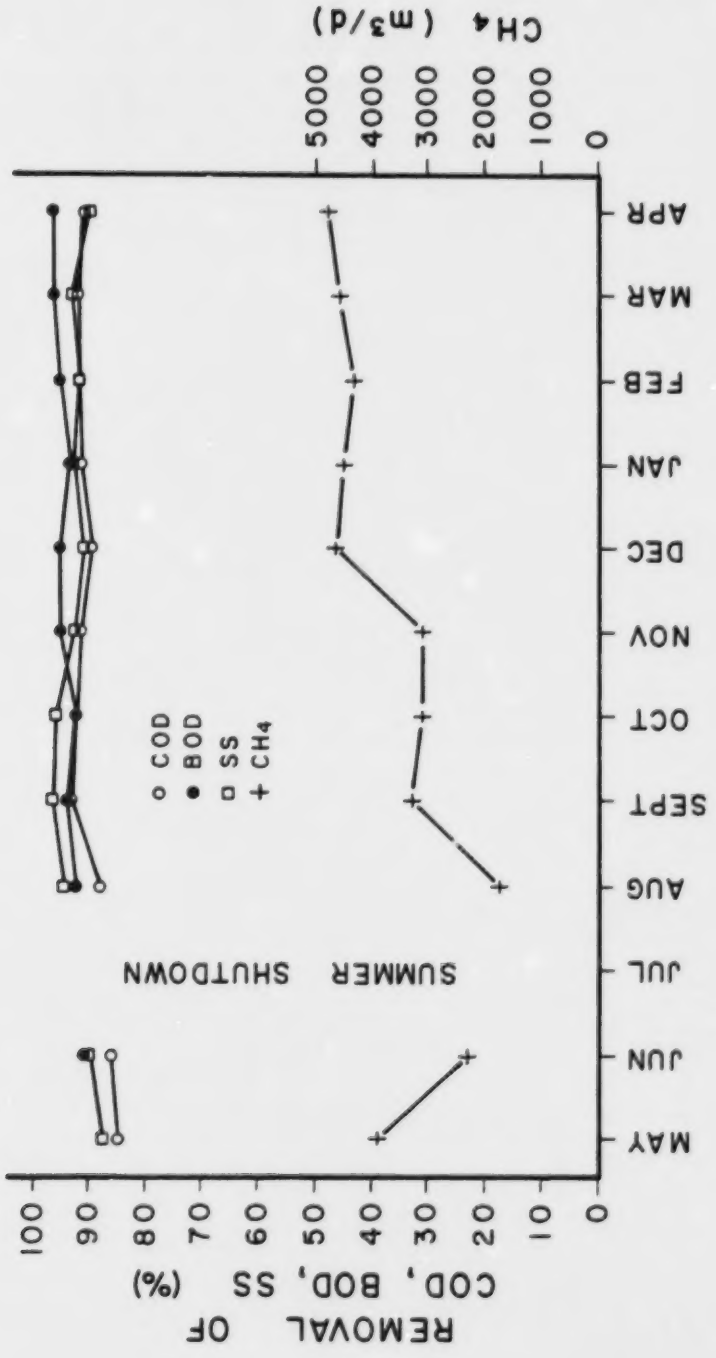

FIG 3: REMOVALS IN ADI-BVF & METHANE

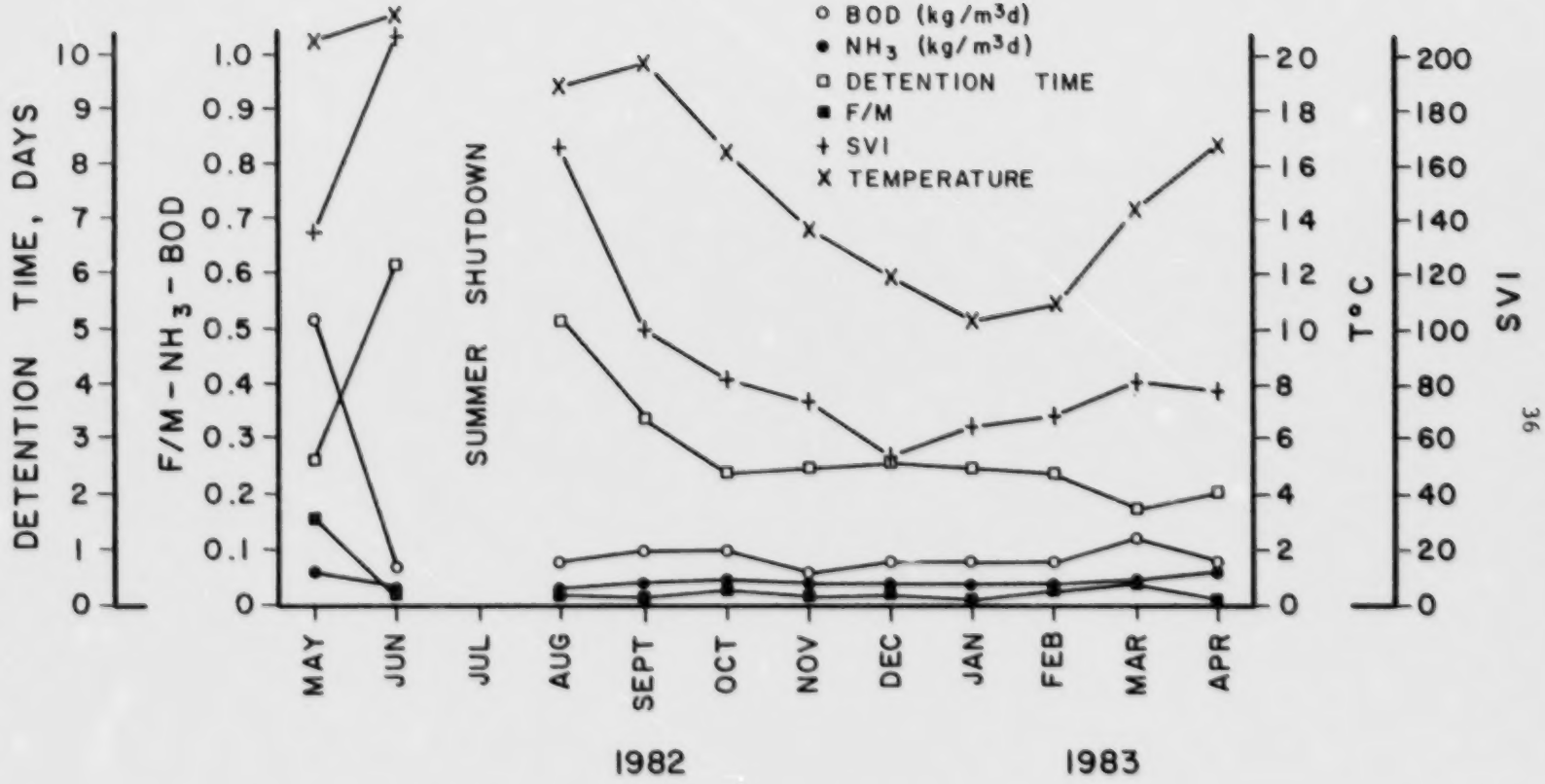

FIG 4: AERATION POND F/M, LOADING, DETENTION TIME AND TEMPERATURE

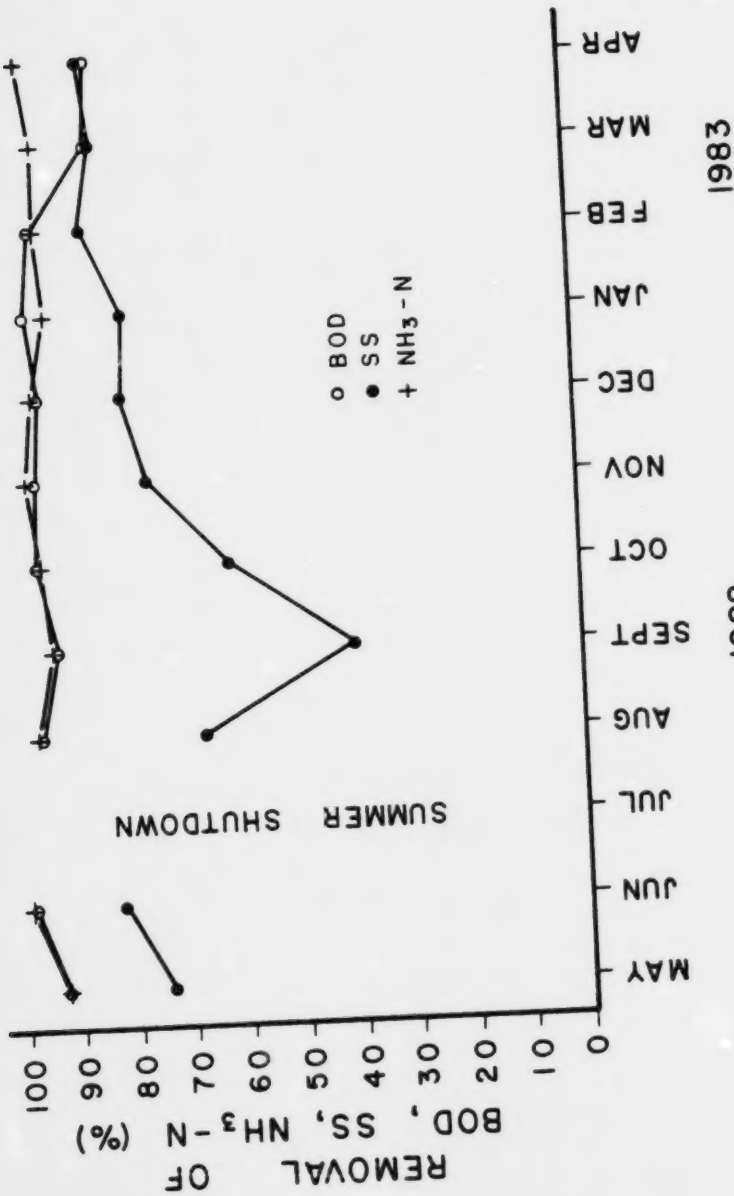

FIG 5: PERFORMANCE OF ACTIVATED SLUDGE

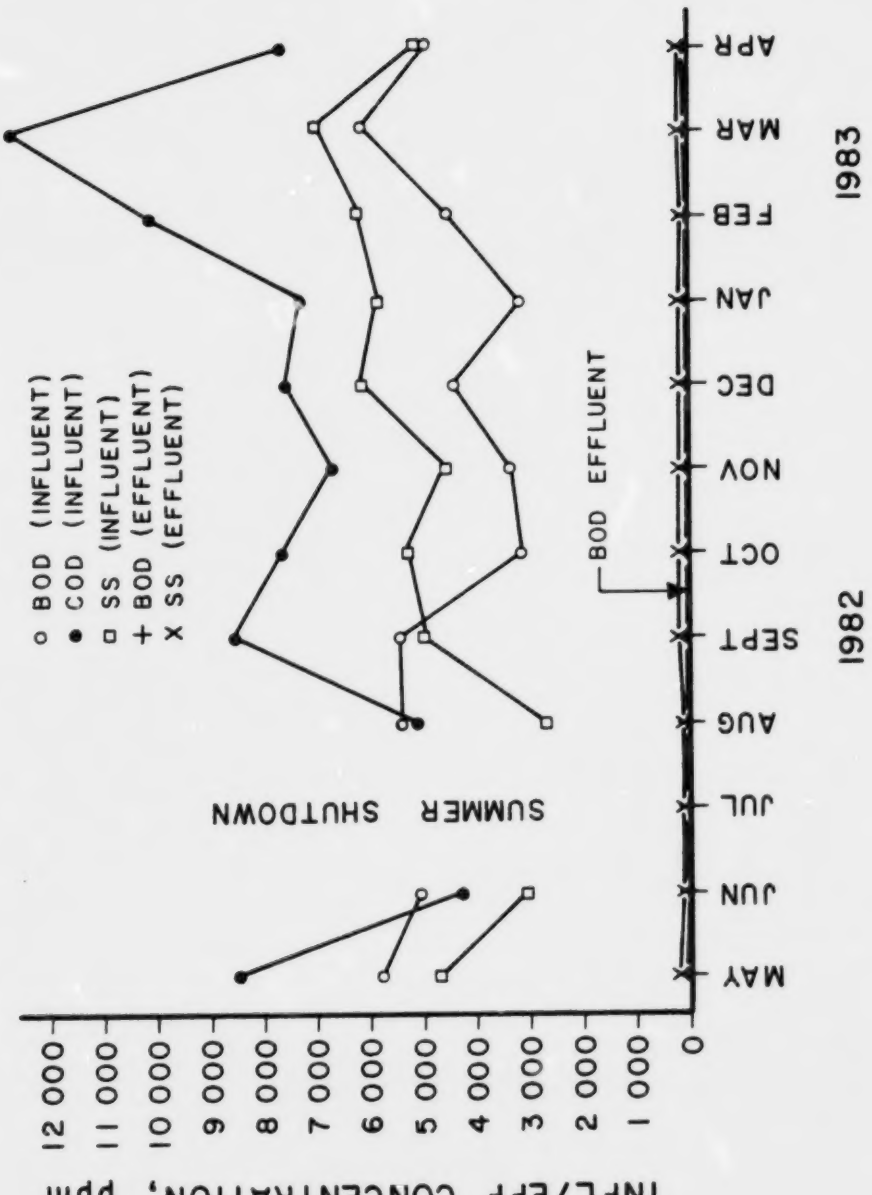

FIG 6: INFLUENT AND FINAL EFFLUENT CONCENTRATIONS

range except for March and April, 1983, when the effluent BOD was 41 and 26 ppm, respectively. The reason for the tailing off in BOD removal during this period is not known for certain at the time of writing, although it is believed to be a related to malfunctioning of the DO pacing equipment together with a change in the sludge wasting schedule.

The SS of the wastewater entering the fermenter was typically in the 4-6 000 ppm range. The final effluent SS was approximately 200 ppm during March-May, 1982, due to difficulties in settleability of the sludge. However, from June, 1982 to April, 1983, it was in the 50-100 ppm range.

From Figure 7 it can be seen that the overall plant removal for BOD, SS and COD was high. Overall COD removal averaged 98%, BOD 99.5% and SS 98%. Even the total nitrogen (sum of all forms) removal was good, being typically better than 80%.

Graphs showing pH, alkalinity and volatile acids in the fermenter appear in Figure 8. It is apparent that the fermenter effluent was very steady near pH 6.5, while the influent pH was typically a little higher. The final effluent pH was also rather steady, typically 7.6-7.8. The alkalinity fell from 1500 ppm in May to approximately 1100 ppm by November, reflecting the reduction in chemicals (alkalinity) fed into the raw waste entering the fermenter, where it has remained. The volatile acids have shown a trend toward decreasing concentration since August, despite increasing loadings and decreasing retention times; minimum VA values were found in January at 35 ppm. Values of 100 ppm or less are considered very low and indicative of a mature sludge, well-acclimatized to the substrate and operating without any appreciable stress.

Discussion

Good performance was achieved in the fermenter even though there were large fluctuations in waste strength from day to day, including values well above nominal design values. Table 3 has been prepared to show the variations in COD and SS entering and leaving the fermenter. Maximum raw COD values were in excess of 20 000 ppm, whereas maximum SS values were in excess of 10 000 ppm. COD and SS removals, were not adversely affected on days or months having high feed strengths. The average SS concentration leaving the fermenter was less than one would expect from a primary clarifier treating potato wastewater.

The maximum and minimum values for certain parameters are furnished in Table 4. The ratio of average COD to BOD ranged from 5:1 to 15:1, indicating a well-stabilized (potato waste) effluent. The ratio of SS to BOD is very high. Setting aside the possibility of anomalous test results, this suggests a stable sludge well into the endogenous respiration phase-supported by the low F/M value, the high solids retention time and the low volatile component of the MLSS, 50-65%.

Average final effluent ammonia values were typically around 5 ppm compared with 100 ppm or more entering the aeration basin. Effluent nitrites were typically under 1 ppm and effluent nitrates typically 15 ppm. Total kjeldahl values in the effluent were usually under 10 ppm. Nitrogen mass balance calculations showed a large loss of nitrogen between the fermenter and the final effluent (after accounting for nitrogen contained in the waste sludge). The only plausible explanation for the nitrogen loss appears to be denitrification. Furthermore, due to the lack of a lot of nitrogen gas

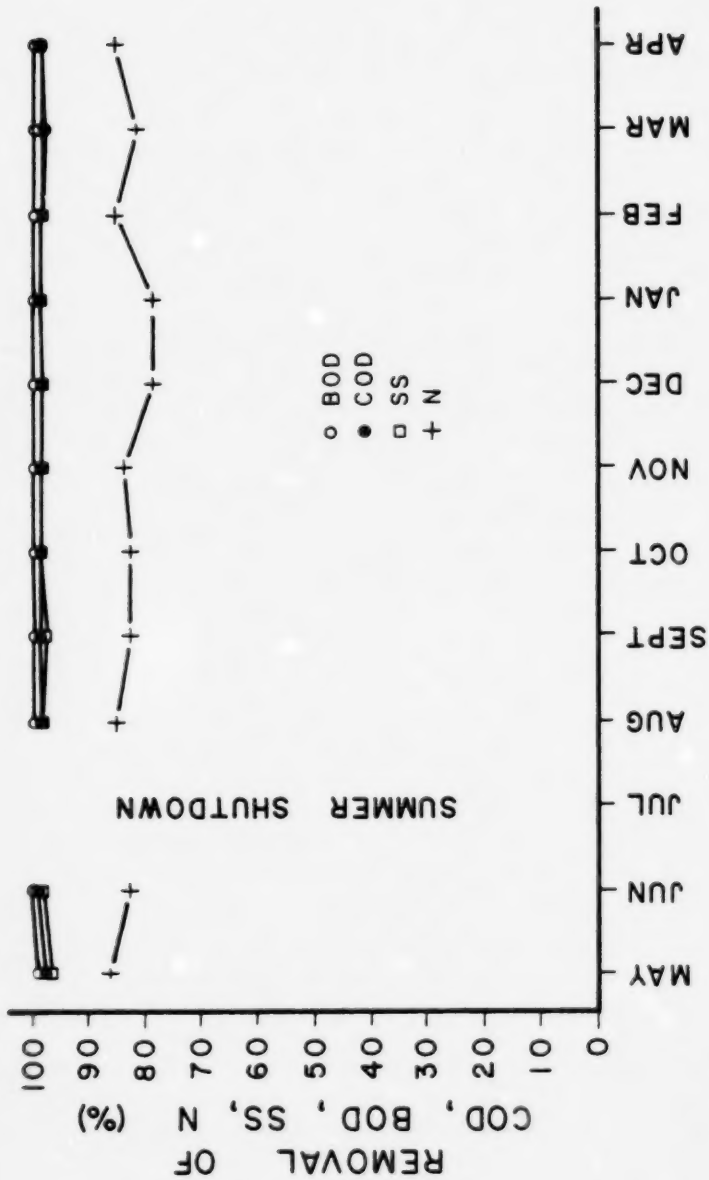

FIG 7: OVERALL REMOVALS

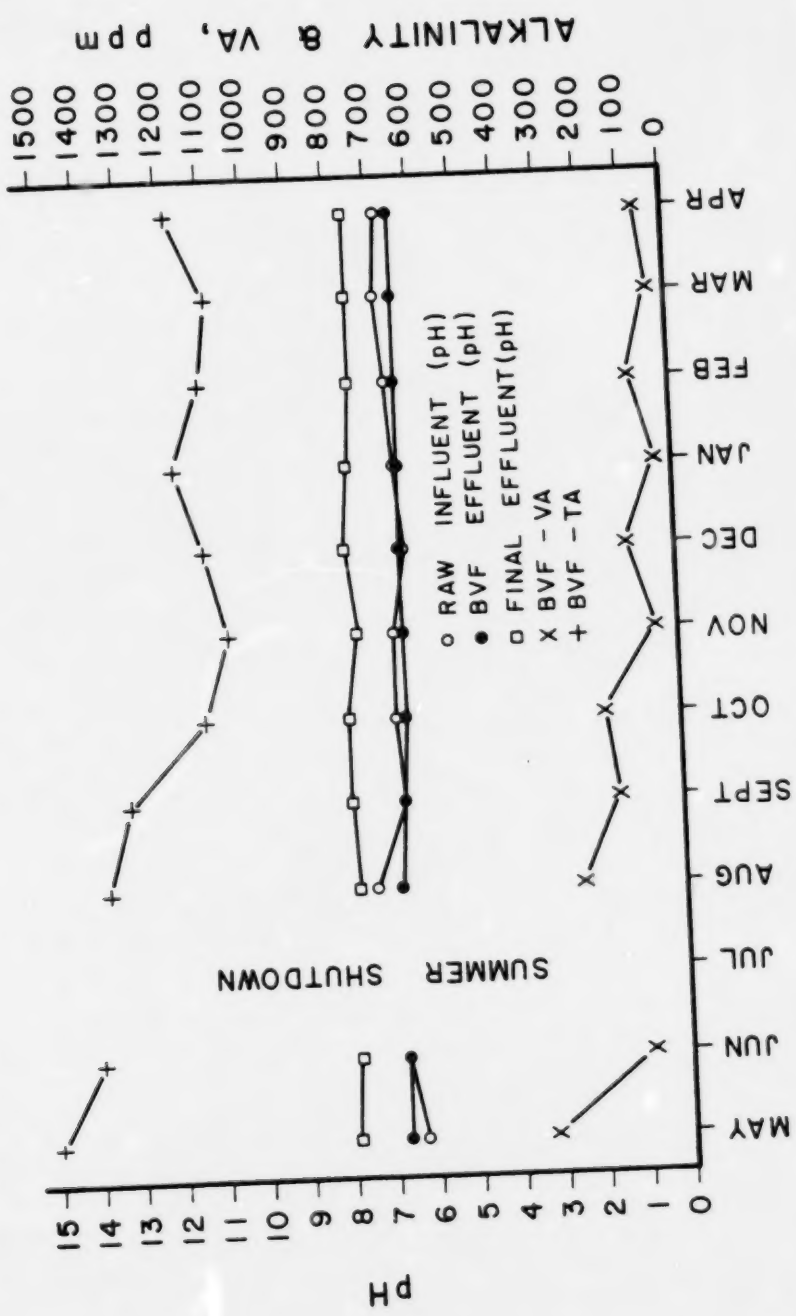

FIG 8: pH, ALKALINITY AND VOLATILE ACIDS

TABLE 3 - MAXIMA AND MINIMA FOR COD AND SS ENTERING AND LEAVING FERMENTER*

	SCREENED INFLUENT						ADI-BVF EFFLUENT					
	COD ppm			SS ppm			COD ppm			SS ppm		
	MAX	MIN	AVG	MAX	MIN	AVG	MAX	MIN	AVG	MAX	MIN	AVG
MAY (1982)	15 500	5 000	8 472	10 550	2 230	4 706	2 100	750	1 281	1 050	315	612
JUNE	9 000	1 000	4 286	5 250	800	3 042	1 000	250	604	505	105	317
JULY	- - - - - - - - - - - - - - - - SUMMER SHUTDOWN - - - - - - - - - - - - - - - -											
AUGUST	9 500	1 250	5 109	5 020	1 020	2 700	1 100	250	613	228	78	161
SEPTEMBER	14 750	4 000	8 534	8 970	1 350	4 994	950	400	581	328	116	193
OCTOBER	23 750	4 250	7 667	12 970	2 740	5 272	825	350	550	336	156	223
NOVEMBER	12 250	2 750	6 702	7 250	1 690	4 558	1 300	300	567	1 088	164	349
DECEMBER	12 250	3 000	7 609	10 780	1 300	6 127	1 200	375	772	1 080	188	572
JANUARY (1983)	10 750	3 500	7 638	9 390	1 460	5 885	1 450	350	645	1 268	168	437
FEBRUARY	20 000	5 000	10 102	11 040	2 620	6 184	2 500	350	856	835	204	489
MARCH	22 500	6 000	12 065	11 750	2 800	7 019	1 900	400	909	1 120	135	514
APRIL	16 000	5 000	7 625	8 570	3 600	5 260	1 250	500	743	835	260	515

* Underlined values indicate overall maximum/minimum in column.

escaping in the final clarifier, the denitrification must have occurred mainly within the aeration basin. And this was possible, as the aerators were turned off and on to save energy; records show that, since fall startup, neither aerator was on for 50-60% of the time, resulting in zero oxygen supplied for 15-30 minute intervals. Denitrification was unplanned, and this turned out to be a pleasant surprise, both to the owner and to the water authority. It, of course, reduced the amount of oxygen and energy input while minimizing the amount of nitrogen compounds entering the receiving water. This phenomenon may not have occurred so fully had it not been for the capability of utilizing the ADI- BVF as storage to discharge a more or less continuous feed supply to the aeration basin seven days per week.

The average monthly COD loading on the fermenter only exceeded design in March, 1983. However, the SS loading exceeded design consistently from September-April, being virtually twice design in March. Yet, the accumulation of potato solids and other sludge has not presented a problem. In fact, according to measurements taken by the operator in January, 1983, there were fewer tons of sludge dry solids present then than there were added as seed sludge during startup, i.e., 1900 m^3 at 3.7% solids, compared with 4400 m^3 at 4% solids of seed sludge added. From startup to January, 1983, the estimated quantity of SS fed to the fermenter was 2-3 million kilograms. Even allowing for possible errors in measurement of accumulated sludge, it is apparent that there is a high degree of destruction of the incoming potato solids. This supports the original decision to delete the primary clarification step, thereby precluding the bother and expense of primary sludge dewatering and disposal while increasing the quantity of biogas energy recovered.

Not only was primary sludge handling circumvented, but dewatering and disposing of secondary sludge was also avoided by wasting it into the ADI-BVF. This further increased gas production, too.

It was expected that some sludge removal from the fermenter would be required every year or so following the initial removal expected after some four years of operation. With the data generated to date, this would appear to be a conservative estimate. In any case, no sludge of any kind--primary, secondary or anaerobic--has been removed or hauled away for disposal to date. This treatment facility, coupled with no need to add any chemicals, nor any heat, and no need to worry about large variations in waste flow and strength, has provided a simple, low-cost system to operate.

Summary and Conclusions

1. The treatment facilities comprised of anaerobic and aerobic treatment successfully treated a food processing wastewater while providing renewable energy in the form of biogas.
2. The anaerobic portion of the plant, an ADI-BVF, provided over 90% removal of BOD, COD and SS while operating under higher loadings than assumed for design and at temperatures lower than 20°C.
3. The ADI-BVF is operating without any addition of chemicals or heat.
4. Primary and secondary sludge are disposed in the ADI-BVF where they are anaerobically decomposed to enhance biogas production. No sludge of any kind has been removed from the facilities after two years of operation, nor is there any evidence of a need to remove sludge in the near future.

TABLE 4 - FINAL EFFLUENT MAX-MIN VALUES FOR COD, BOD, SS, AMMONIA AND NITRATES*

	COD, ppm			BOD, ppm			SS, ppm			NH_3-N, ppm			NO_3, ppm		
	MAX	MIN	AVG	MAX	MIN	AVG	MAX	MIN	AVG	MAX	MIN	AVG	MAX	MIN	AVG
MAY (1982)	420	40	197	115	17	64	336	44	159	32	2	12	13	4	10
JUNE	90	15	38	14	3	8	104	32	56	8	0	3	31	19	26
JULY	- - -

5. Biogas is recovered and used in the factory boilers to make process steam.
6. The activated sludge plant removed over 90% of the BOD and ammonia.
7. Overall removals were: COD 98%, BOD 99.5%, SS 98% and total nitrogen 80%. High nitrogen removals were attributed to denitrification in the aeration plant.
8. The entire treatment facility was economical to construct and simple to operate, requiring only two operators. The monetary value of the biogas recovered is greater than all the operating costs including labor, chemicals, electricity, maintenance, discharge tax and sundries.

DESIGN OF A HIGH TECH WASTEWATER TREATMENT PLANT IN THE PULP AND PAPER INDUSTRY

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PERFORMANCE OF WASTEWATER TREATMENT SYSTEMS IN THE U.S. PULP' PAPER AND PAPERBOARD INDUSTRY

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LE TRAITEMENT PHYSICO-CHIMIQUE DES EFFLUENTS URBAINS PAR UTILISATION CONJOINTE D'UN SEL FERREUX, D'AIR ET D'UN SEL FERRIQUE

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RÉSUMÉ

Différents réactifs ont été testés, seuls ou en association avec d'autres, sur une installation existantes de traitement physico-chimique d'effluents urbains.

L'association d'un sel ferreux, d'air et d'un sel ferrique s'est révélée être la combinaison la plus performante sur le plan de la souplesse et de la sécurité d'exploitation, les coûts de fonctionnement étant inchangés par rapport à ceux des solutions classiques. En particulier, les risques de dégagement de mauvaises odeurs se sont trouvés complètement maîtrisés.

Le procédé permet d'utiliser les sels de fer ferreux sous produits par l'industrie (décaperie, fabrication de l'oxyde de titane) et induit par conséquent une dépollution étendue.

À cette valorisation, viennent se conjuguer une régénération et un recyclage partiel des réactifs usés à l'intérieur même de l'usine d'épuration. Des réactions biologiques insoupçonnables jusqu'alors sur des usines d'épuration physico-chimiques, sont à l'origine de ce recyclage partiel dont le coût est dérisoire.

Moyennant certains aménagements, le procédé peut être appliqué à des usines d'épuration existantes, dont on souhaite améliorer les résultats.

LE TRAITEMENT PHYSICO-CHIMIQUE DES EFFLUENTS URBAINS PAR UTILISATION CONJOINTE D'UN SEL FERREUX, D'AIR ET D'UN SEL FERRIQUE

L'inertie des processus biologiques d'épuration les rend inaptes à assurer le traitement convenable des eaux résiduaires urbaines soumises à des variations importantes de leurs caractéristiques. C'est le cas notamment des régions touristiques sujettes à des fluctuations saisonnières brutales de population. En raison de leur souplesse, les procédés physico-chimiques ont pris le relais des stations biologiques, ou ont été couplés avec elles.

L'application inconsidérée aux effluents urbains de réactifs utilisés pour les traitements physico-chimiques traditionnels a conduit à des échecs cuisants, tant sur le plan des résultats techniques que sur celui des risques sanitaires et de l'économie. De la gamme des produits disponibles sur le marché, il convient par conséquent de retenir ceux qui assurent à la fois la stabilité des performances, la sécurité de fonctionnement et l'économie d'exploitation.

L'objet du présent article est de présenter les possibilités offertes par l'utilisation conjointe de sels ferreux, d'air et de sels ferriques pour la floculation des eaux.

Cette association de réactifs a été testée sur une station d'épuration existante, en zone touristique, d'une capacité de 45 000 équivalents habitants et prévue initialement pour fonctionner au sulfate d'alumine et à la chaux. Les essais ont duré pendant les 2 mois de service annuel de cette station, et qui correspondent précisément à la période de pointe de rejets urbains. La comparaison avec les performances obtenues lors de la saison estivale précédente, où le sulfate d'alumine et la chaux ont été utilisés, a apporté des enseignements intéressants.

Caractéristiques de l'usine d'épuration physico-chimique:

-	population raccordée	45 000 équivalents-habitants
-	dotation hydrique par équivalent habitant	150 l/j (*)
-	charge polluante par équivalent-habitant DBO_5	60 g/j (*)
-	volume journalier d'effluents à traiter	6 750 m^3
-	DBO_5 journalière à traiter	2 700 kg
-	DCO journalière à traiter	5 400 kg
-	MES journalière à traiter	2 700 kg

Bien que cette usine n'ait pas été destinée à recevoir des matières de vidanges, des rejets de ces substances y parvenaient clandestinement par le biais des égouts.

En période hivernale, l'usine physico-chimique n'est pas en service. Le traitement des effluents rejetés par la population sédentaire s'effectue dans une petite usine d'épuration biologique, totalement indépendante, tout au long de l'année.

Le coefficient de pointe des rejets à l'égout entre l'hiver et l'été s'élève à 9.

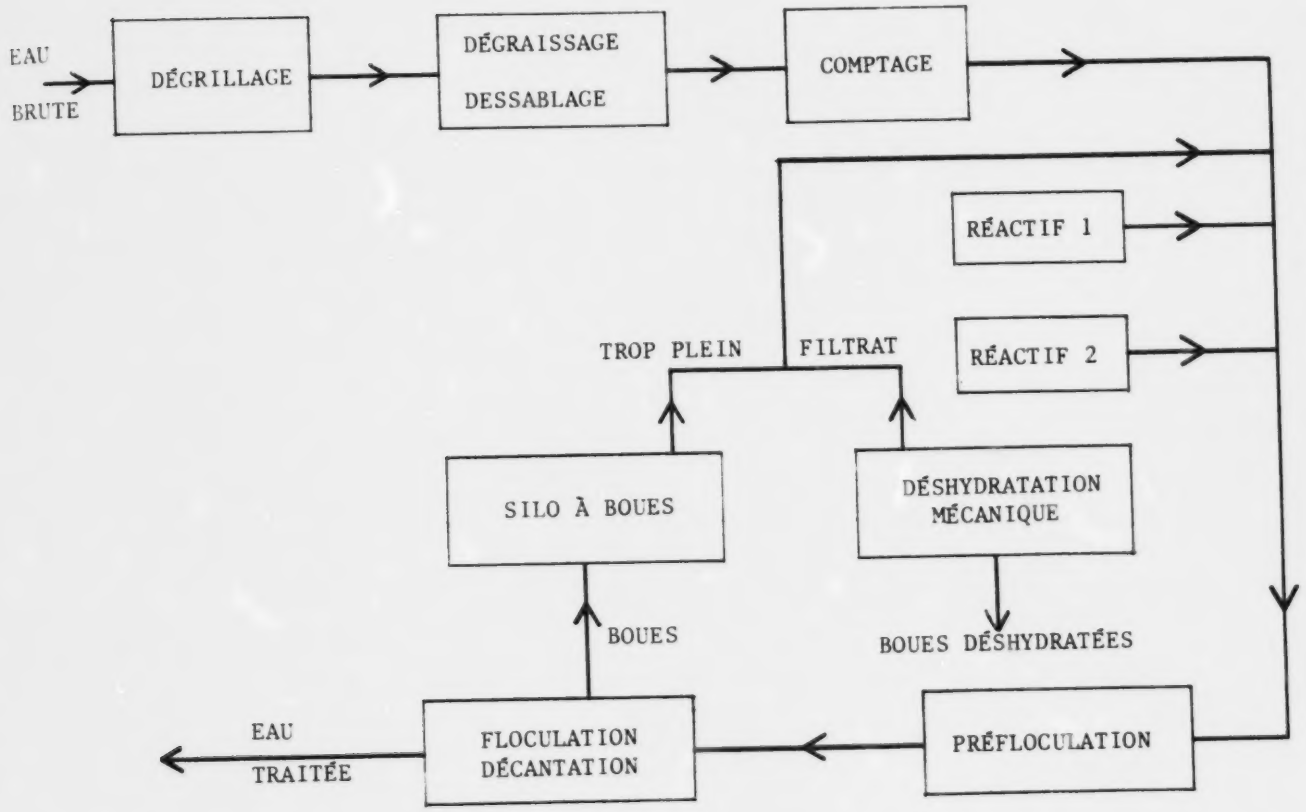
La filière de traitement de l'usine d'épuration physico-chimique, telle qu'elle était construite au départ, est la suivante: (fig. 1)

- un dégrillage automatique;
- un bassin de prétraitement combiné, assurant à la fois un dégraissage par insufflation d'air (moyenne bulle) et un dessablage automatique avec lavage et essorage des sables;
- un canal de comptage des effluents comportant un débitmètre "bulle à bulle", régulant l'injection de réactifs;
- un préfloculateur brassé par un hélico-mélangeur ou par de l'air prévu en secours;
- un silo à boues;
- une déshydratation mécanique des boues par filtre à bandes presseuses;
- un stockage de réactifs: . un silo pour la chaux
  . des citernes pour le sulfate d'alumine

(*) Il s'agit là de chiffres standard moyens utilisés dans le cadre du projet. Dans la réalité, les dotations en charge polluante et en débit par équivalent-habitant sont inférieures en zone touristique. Ceci permet d'expliquer que l'usine décrite dans la présente étude et prévue pour 45 000 habitants, ait pu faire face à une population présente sur le site avoisinant 100 000 habitants.

FIGURE 1

SCHÉMA DE PRINCIPE DE L'INSTALLATION



INSUFFISANCES DE LA FILIÈRE DE TRAITEMENT AU SULFATE D'ALUMINE ET À LA CHAUX

Malgré des performances intéressantes sur le plan de l'élimination des matières organiques, cette filière s'est exclue d'elle-même à cause de déboires liés aux odeurs.

Des odeurs intolérables se sont dégagées de la surface libre des plans d'eau par suite du dégazage spontané de composés soufrés nauséabonds présents dans l'effluent en cours de traitement. Dans le silo concentrateur, où les putréfactions sont intenses, les dégagements d'odeurs étaient particulièrement importants. Il a fallu injecter massivement de la chaux pour bloquer les sulfures responsables de ces émanations. La contrepartie de cet ajout de chaux a été le dégagement d'ammoniac, qui a obligé à recourir parfois au masque à gaz dans l'atelier de traitement des boues. Les circuits en cuivre des armoires électriques ont eu à souffrir de ces émanations ammoniacales.

La sécurité d'exploitation n'ayant pas pu être obtenue dans des conditions satisfaisantes, la filière au sulfate d'alumine et à la chaux a dû être abandonnée.

LA FILIÈRE AUX SELS FERREUX ET FERRIQUE

Caractéristiques des nouveaux réactifs utilisés

Le sulfate ferreux en cristaux heptahydratés est un sous-produit des industries de transformation: tréfileries, laminoirs de tôles, etc.

Celui qui a été choisi pour les essais est un coproduit de la fabrication du bioxyde de titane.

Ses caractéristiques chimiques sont les suivantes:
(toutes les données sont en poids par poids)

- $FeSO_4, 7H_2O$ = 93 - 95%
- Fe = 18,5%
- TiO_2 = 1 500 - 2 000 ppm
- Mn = 2 000 ppm
- Cr = 1,5 ppm
- Sb = 5 ppm
- Ni = 15 ppm
- V = 8 ppm
- Cu = 0,5 ppm
- Pb = 1 ppm
- Zn = 150 ppm
- Co = 30 ppm
- Cd = 0,2 ppm

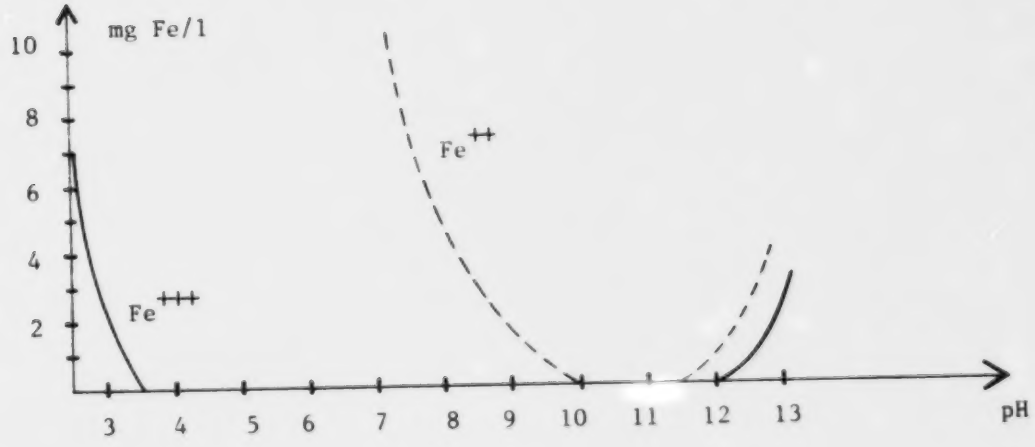
Au chlorure ferrique traditionnel, il a été préféré le chlorosulfate ferrique, obtenu par action du chlore sur le sulfate ferreux décrit précédemment; l'emploi du chlorosulfate ferrique répond ici au souci de revalorisation des déchets industriels.

Le chlorosulfate ferrique commercial se présente sous forme d'une solution concentrée de densité 1,48 à 20°C, et contenant 180 g de fer par litre.

Les courbes de précipitations des hydroxydes ferreux et ferrique apparaissent à la figure 2.

FIGURE 2

COURBES DE PRÉCIPITATION DES HYDROXYDES FERREUX ET FERRIQUE



Préalablement aux essais sur l'installation réelle, des essais ont été effectués au jar test.

Les réactifs ci-après ont été utilisés:

- sulfate ferreux seul;
- sulfate ferreux suivi d'une aération;
- chlorosulfate ferrique seul;
- sulfate ferreux suivi d'une aération puis d'une injection de chlorosulfate ferrique;
- sulfate d'alumine seul, à titre de comparaison.

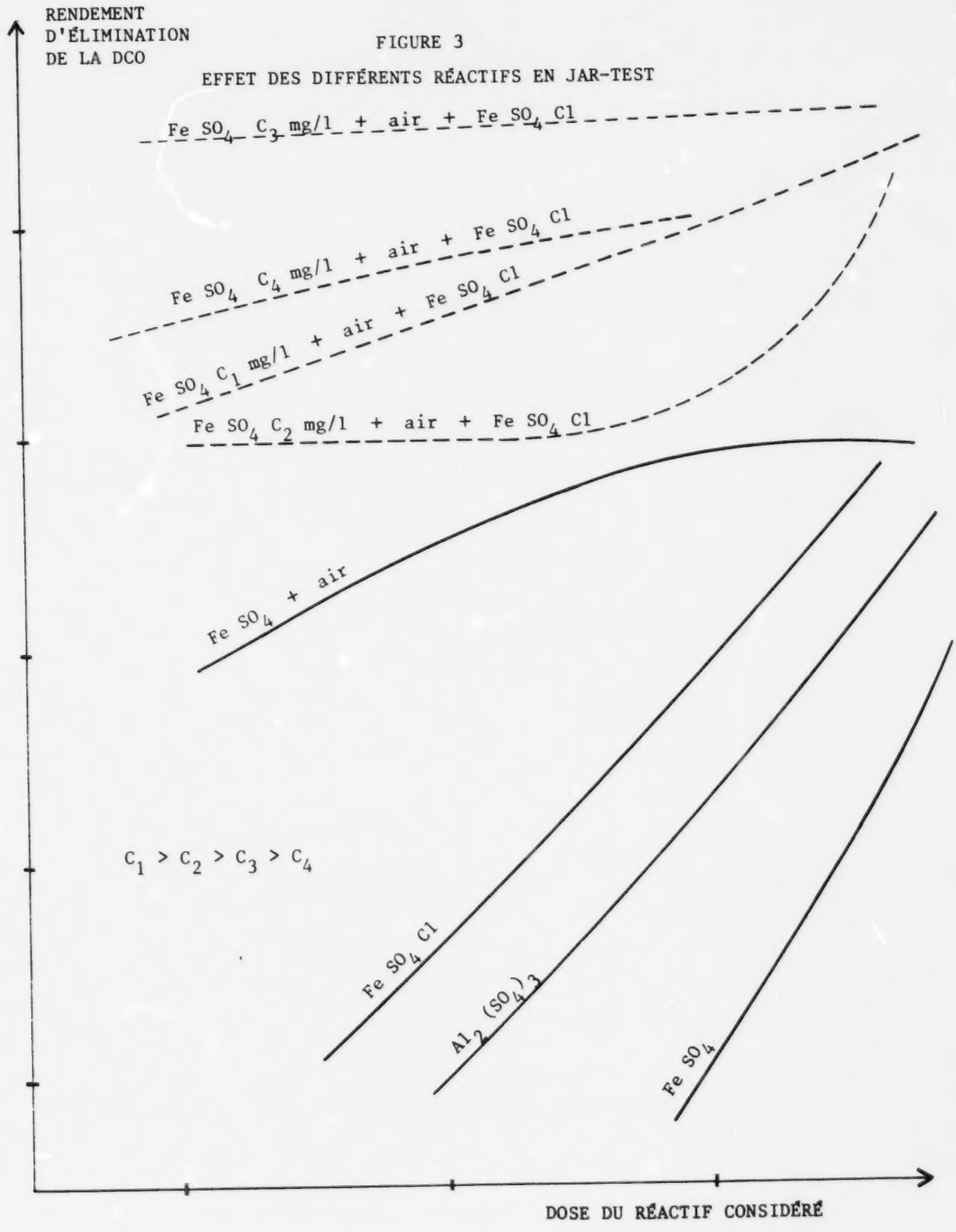
En raison des "effets de bord" inhérents aux appareils de laboratoire, la détermination des MES de l'eau traitée n'a pas été effectuée. La mesure de la DCO, plus significative, lui a été préférée.

La figure 3 de la page 8 fait la synthèse des résultats obtenus à l'issue de ces essais.

Le graphique de cette figure montre que l'association fer ferreux, air, fer ferrique, est la plus performante de tous les réactifs testés: elle se traduit par un effet synergique des deux formes de fer.

FIGURE 3

EFFET DES DIFFÉRENTS RÉACTIFS EN JAR-TEST



Cette synergie ne se manifeste pas seulement sur la qualité de l'eau épurée, mais également sur l'aptitude à la décantation du floc formé.

L'idée de l'utilisation conjointe du fer ferreux et du fer ferrique repose sur l'observation, fréquente en eaux industrielles, que les complexes ferroso-ferriques α (FeO,nH_2O) β (Fe_2O_3, $n'H_2O$) avec $\alpha + \beta = 1$, présentent une meilleure aptitude à la floculation et à la décantation que les flocs ferreux et les flocs ferriques séparés.

Les essais de floculation en jar test, qui ont conduit au graphique de la figure 3, et décrits ci-après, permettent de comprendre ce qui se passe lors de l'injection des réactifs.

À de l'effluent brut, modérément brassé par de l'air, on ajoute progressivement du sulfate ferreux. Une coloration noire se développe, qui s'intensifie avec l'ajout du réactif. Cette coloration noire est due à la formation de sulfure ferreux.

$$Fe^{++} + S^{--} \longrightarrow \underline{FeS}\searrow$$

qui "empoisonne" la floculo-décantation de l'hydroxyde ferreux formé conjointement. Afin de ne pas trop dilacérer ce floc fragile, l'aération a dû être arrêtée assez vite. De ce fait, l'oxydation des composés ferreux, et par suite l'éclaircissement de la couleur du milieu, ont été peu marqués ($\alpha \simeq 1$, $\beta \simeq 0$).

L'ajout de chlorosulfate ferrique auquel on procède ensuite, doit normalement donner naissance à des dépôts rouille classiques, qui devraient, par superposition de couleur, éclaircir le précipité noir précédemment formé. Or il n'en est rien, la couleur noire ne fait que se renforcer, tout au moins lorsqu'une quantité modérée de fer ferrique a été ajoutée.

Cette intensification du noircissement est due à une nouvelle formation de sulfure de fer dans le milieu.

(1) $2\,Fe^{+++} + S^{--} \longrightarrow 2\,Fe^{++} + \underline{S}\searrow$

(oxydant)

puis, si des sulfures persistent dans le milieu

(2) $Fe^{++} + S^{--} \longrightarrow \underline{FeS}\searrow$

noir, inodore

soit, au total

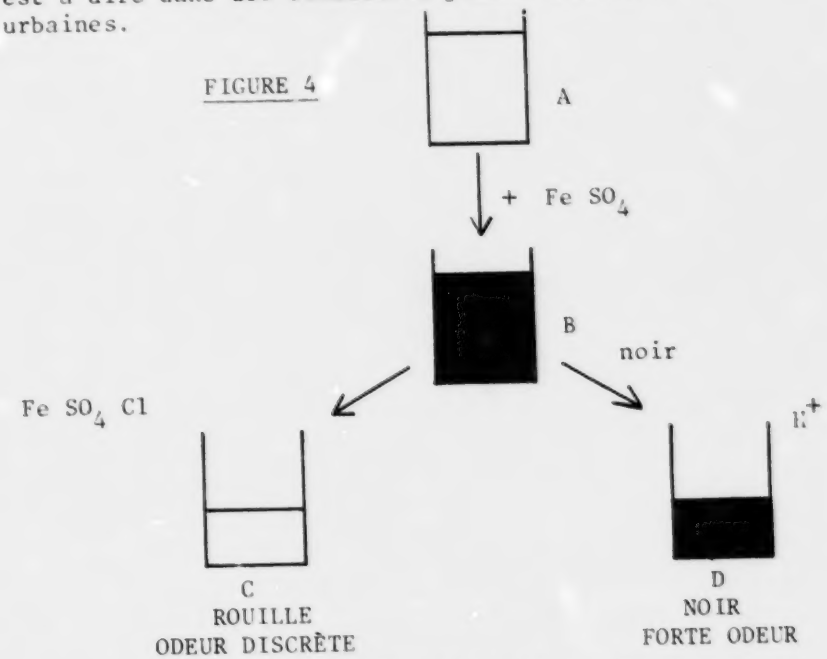
(3) $2\,Fe^{+++} + 3\,S^{--} \longrightarrow 2\,\underline{FeS}\searrow + \underline{S}\searrow$

Si l'on poursuit l'ajout de chlorosulfate ferrique une fois que tous les ions S^{--} ont été consommés, le floc rouille du sel ferrique n'apparaît pas comme on pourrait s'y attendre. Lorsque l'ajout de chlorosulfate ferrique est suffisant, le floc est d'un brun clair. La couleur noire du sulfure de fer a disparu et n'est pas masquée par le floc ferrique.

On pourrait penser que la disparition du sulfure de fer provient de l'attaque de ce sulfure par l'acidité libérée lors de l'hydrolyse du chlorosulfate ferrique. L'expérience suivante prouve qu'il n'en est rien:

À un volume donné d'effluent brut (récipient A), on ajoute du sulfate ferreux qui conduit à un noircissement du milieu (récipient B). On divise le contenu du récipient B en deux parties égales. La première partie (récipient C) reçoit du chlorosulfate ferrique en quantité telle que la couleur noire du milieu disparaît, laissant la place à une couleur rouille. La seconde fraction (récipient D) reçoit un acide fort en quantité égale à celle libérée dans le récipient C. Aucune décoloration ne s'y produit. Il faut ajouter 20 à 30 fois plus d'acide fort pour que la disparition du noir ait lieu, c'est à dire dans des conditions jamais atteintes en traitement des eaux urbaines.

FIGURE 4



La conclusion de cette expérience est que l'acidité n'est pas la cause de la disparition de la couleur noire. Au reste, en fin d'expérience, le récipient D émet des odeurs putrides renforçées par l'acidité, tandis que l'eau du récipient C voit son odeur diminuer malgré une acidification du milieu. La disparition de la couleur noire sous l'effet du chlorosulfate ferrique est due au pouvoir oxydant de ce réactif vis-à-vis du sulfure de fer:

(4) $2\ Fe^{+++} + FeS \longrightarrow 3\ Fe^{++} + S \searrow$

À ce stade, et malgré l'ajout de fer ferrique, il n'existe dans le milieu que du fer bivalent ($\alpha = 1$, $\beta = 0$).

Cette formation de fer bivalent dans le milieu a été confirmée par l'expérience qui suit:

À un volume donné d'effluent brut, on ajoute 800 mg/l de $FeSO_4$, $7H_2O$, correspondant à 150 mg Fe/l. Après floculation décantation, la dose de fer ferreux libre que l'on retrouve dans l'eau surnageante est négligeable.

À un même volume d'une autre fraction d'effluent brut, on ajoute 1 250 mg/l de solution commerciale de chlorosulfate ferrique, ce qui correspond au même ajout de fer que dans l'expérience précédente (150 mg Fe/l). Après floculation décantation, l'eau surnageante contient 10 ppm environ de Fe^{++}, confirmant que l'équation chimique (4) a bien lieu dans le contexte physico-chimique d'un effluent urbain.

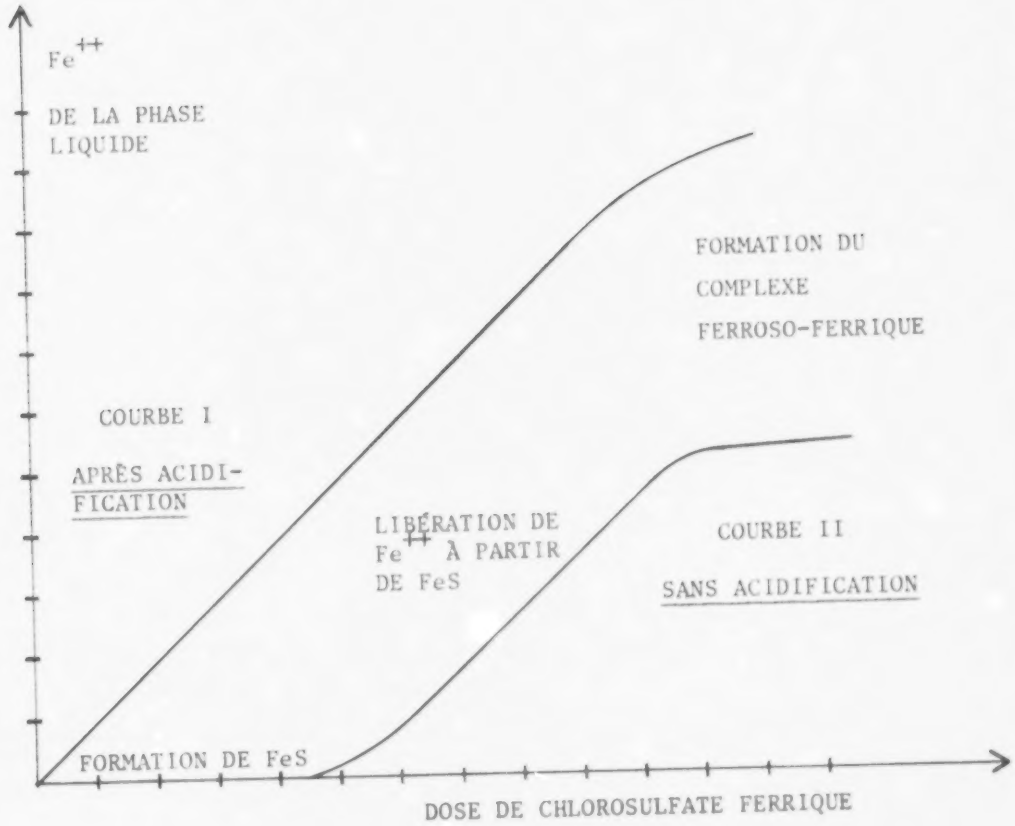
Les 140 mg de fer restant se retrouvent dans le fond du récipient sous forme de précipité ferroso-ferrique.

Afin de déterminer α et β (du complexe ferroso-ferrique), on acidifie fortement le milieu. Le sulfure de fer, l'hydroxyde ferreux, l'hydroxyde ferrique se solubilisent. La forte acidité du milieu (ph < 1) inhibe l'oxydation par l'air du fer ferreux. De ce fait, il n'est plus nécessaire de manipuler les échantillons sous azote.

Les graphiques de la figure 5 rendent compte de l'évolution de la teneur en fer ferreux dissous d'un effluent urbain ayant reçu une dose croissante de chlorosulfate ferrique.

FIGURE 5

ÉVOLUTION DE LA TENEUR EN FER FERREUX
EN FONCTION DE L'AJOUT DE SEL FERRIQUE



L'idée de l'utilisation préalable du sulfate ferreux reposait sur l'intention de bloquer les sulfures à vil prix et de réserver le chlorosulfate ferrique pour la floculation à proprement parler. Or nous avons vu que le sulfure de fer formé à partir du sulfate ferreux servira lui aussi à réduire le chlorosulfate ferrique. Autrement dit, qu'il y ait ou non action préalable du sulfate ferreux, le chlorosulfate ferrique aura à oxyder les sulfures - sous quelque forme qu'ils soient - à l'état de soufre.

On peut se demander dès lors où réside l'intérêt d'une utilisation conjointe de sulfate ferreux et de chlorosulfate ferrique. Cet intérêt réside:

- dans les possibilités d'utiliser l'air pour l'oxydation partielle du fer ferreux qui se forme comme le tiers produit au cours de l'action des deux réactifs injectés. L'action de l'air dans le cas de l'utilisation du sulfate ferreux seul ne saurait conduire à un résultat similaire car la vitesse d'oxydation de FeS en milieu aqueux est bien plus lente que celle de Fe^{++}. Une aération trop prolongée de la liqueur ne saurait être effectuée sans briser la cohésion du floc que l'on chercherait à former;

- le sulfure de fer persisterait dans le milieu et empoisonnerait la floculation décantation. L'effluent traité serait noir;

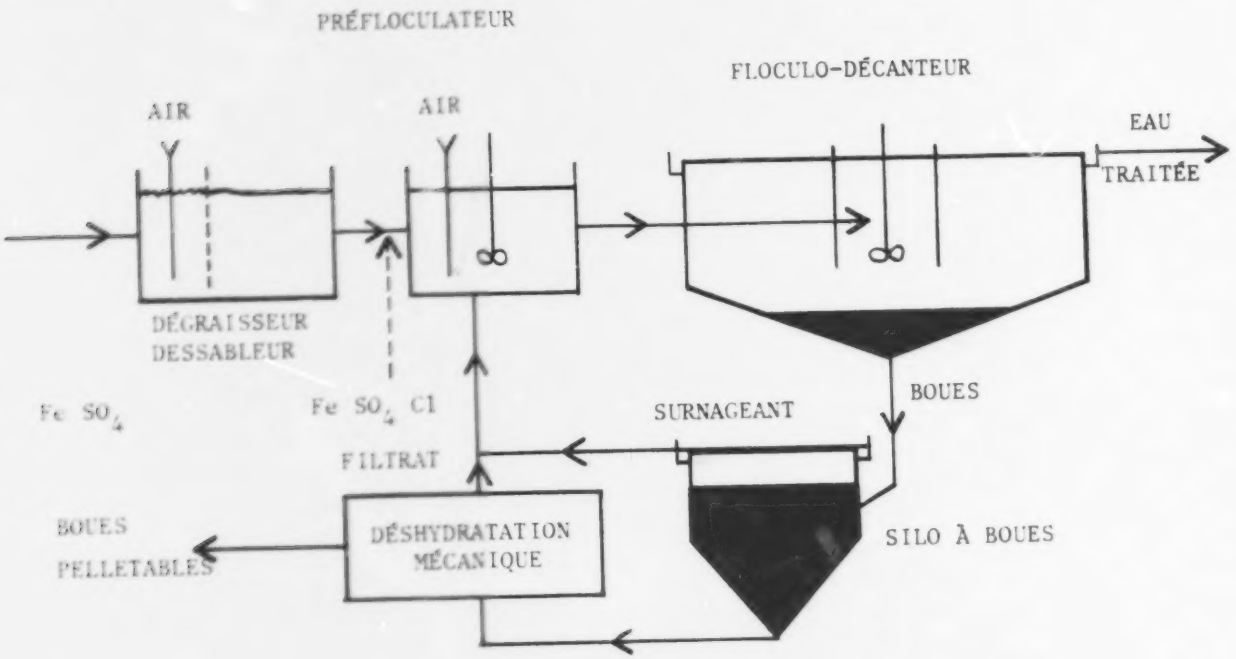
- dans la possibilité de former in situ un floc volumineux et dense, le complexe ferroso-ferrique, qui n'est pas gêné par FeS, détruit.

Remarquons que le soufre qui apparaît à l'état d'élément au cours de ces réactions, n'est pas colloïdal et n'empoisonne donc ni la formation, ni la sédimentation du complexe ferroso-ferrique.

Application industrielle des essais

Les injections de sulfate ferreux et de chlorosulfate ferrique ont été effectuées sur l'installation existante conformément au schéma de principe ci-dessous:

FIGURE 6



Sur ce schéma, l'air est injecté en deux points. En tête, le brassage du dégraisseur dessableur assure l'oxydation d'une partie du fer ferreux apporté en tant que réactif. Un peu plus loin, le préfloculateur oxyde à la fois le fer en provenance du trop plein du silo, qui est un véritable générateur de fer ferreux, et le fer provenant du filtrat de désydratation mécanique des boues. Le recyclage de ce réactif ferreux fabriqué in situ a permis une économie de réactif de l'ordre de 30% par rapport à la consommation de sulfate ferreux telle qu'elle était à la mise en route de l'usine, avant que ces retours n'existent.

La concentration en Fe^{++} du surnageant du silo - 50 à 100 ppm - demeure invariante si on accélère la lixiviation des boues du silo en augmentant le débit des purges du décanteur.

Autrement dit, le flux de fer ferreux recyclé peut augmenter. Une limite est cependant imposée par le régime hydraulique qu'il convient de ne pas dépasser pour le silo.

La concentration en fer ferreux des boues soutirées du décanteur est insuffisante pour expliquer la richesse en fer ferreux à un tel degré du trop plein du silo. C'est l'évolution biologique du silo qui est à l'origine de cette "fabrication" de fer ferreux. Les matières organiques des boues libèrent en effet par fermentation des sulfures qui sont, en cascade, transformés en FeS et finalement en soufre et fer ferreux sous l'effet des boues ferriques existantes. La génération de fer ferreux dans le silo est donc le tiers produit utile de la désodorisation.

L'absence d'odeur à la surface du silo demeure tant qu'il y existe du fer ferrique "antiseptique". Il convient donc de limiter l'âge de la boue en dimensionnant bien ce silo pour qu'il ne diverge pas vers un état totalement réducteur, où les odeurs se dégageraient alors abondamment.

La limite que l'on est contraint de donner à la capacité du silo limite également la récupération du fer ferreux par rinçage du silo à l'aide des purges du décanteur (*).

Il ressort de l'ensemble des considérations précédentes, que des phénomènes biologiques aussi importants qu'inattendus dominent la bonne marche d'une installation physico-chimique.

(*) Remarquons que le recyclage du fer ferreux du silo est accompagné d'un recyclage conjoint de matières organiques dissoutes ou colloïdales. Globalement, le rendement épuratoire de la station n'est affecté que de quelques % sur le plan de la DCO. Le rendement d'élimination des matières en suspension n'est pas, bien entendu, affecté par ce recyclage.

Les performances de la filière de traitement, correspondant à la figure 6, sont indiquées dans le tableau ci-dessous qui fait apparaître, à titre de comparaison, les résultats obtenus avec le sulfate d'alumine.

	SULFATE FERREUX + CHLOROSULFATE FERRIQUE	SULFATE D'ALUMINE
Dose de polyélectrolyte	0 @ 1,5 mg/l	1 @ 2 mg/l
Rendement d'élimination de la DCO	68 @ 73%	55 @ 62%
Rendement d'élimination des MES	89 @ 93%	80 @ 87%
Production de boues Matières sèches/kg DCO éliminée	0,81 kg/kg	0,62 kg/kg
Siccité des boues déshydratées	29 - 34%	24 - 27%

À coût d'exploitation égal, le procédé aux sels ferreux et ferrique se révèle plus performant sur le plan épuratoire. Sur le plan de l'exploitation, la sécurité est meilleure au niveau des deux organes principaux: le décanteur et l'appareil de déshydratation mécanique des boues.

Le floc ferroso-ferrique décante bien, au point de rendre parfois inutile l'injection d'adjuvant organique de floculation. Cette aptitude se retrouve au niveau du traitement des boues, où une siccité bien plus élevée a pu être obtenue.

Cette compacité du volume de boues à évacuer compense largement une production légèrement accrue de ces boues, et qui est la contrepartie d'une meilleure épuration de l'eau.

Diverses combinaisons de réactifs ont été testées sur l'installation réelle d'épuration, afin de voir si les performances relatives du procédé ferroso-ferrique s'avéraient conservées. Les résultats pratiques de ces essais sont les suivants:

1° Installation fonctionnant au sulfate d'alumine seul

La maîtrise des odeurs y est impossible, car les sels d'aluminium sont indifférents vis-à-vis des sulfures.

2° Installation fonctionnant au sulfate ferreux et au sulfate d'alumine

Le blocage des sulfures préexistants s'y opère, mais le sulfure ferreux formé reste en l'état, car le sulfate d'alumine est inactif vis-à-vis de lui. Colloïdal en partie, ce sulfure de fer colore les effluents "traités" en noir. La coloration est plus spectaculaire que dangereuse et l'installation pourrait - à la limite - être jugée satisfaisante, car cette coloration ne recèle pas beaucoup de DCO et il n'y a pas d'odeur au niveau du traitement de l'eau à proprement parler.
C'est au niveau des boues et de leur traitement que les mauvaises odeurs apparaissent. En effet, toute putréfaction sulfureuse des matières organiques n'est ni bloquée ni oxydée.
Pour bloquer ces sulfures, il faut augmenter la dose de sulfate ferreux, mais alors la teneur en fer ferreux de l'eau traitée augmente exagérément.

La stabilisation par l'air des sulfures naissants dans le silo est impossible car l'azote entraîne par stripping des gaz sulfurés vers l'atmosphère. Aussi, certaines unités ont-elles dû recourir à l'oxygène pur pour maîtriser les odeurs.

3° Installation fonctionnant au chlorosulfate ferrique seul

Toutes les installations mettant en oeuvre le chlorosulfate ferrique seul, utilisent toujours ce réactif en surdosage, en regard de l'objectif de floculation qu'on attend de lui. Une partie de ce réactif sert, comme nous l'avons vu, à des réactions d'oxydo-réduction et de précipitation pures et simples, qui n'ont rien à voir avec une véritable floculation.
Pour les raisons que nous avons vues, la fuite en fer ferreux dissous est inévitable dans l'eau décantée, qui peut prendre une couleur rouille. Ce fer ferreux doit être regardé comme une matière en suspension potentielle.

Comme sur toute installation physico-chimique, des économies de réactif ont été recherchées, qui consistent à proportionnaliser l'injection du réactif, non plus au débit, mais à la pollution influente.
La détection en continu par sonde des matières en suspension des eaux brutes, donne généralement un reflet suffisant de la pollution entrante et présente un temps de réponse compatible avec les exigences du traitement physico-chimique. L'asservissement de l'injection du chlorosulfate ferrique au taux de matières en suspension, ne s'est pas avéré ici possible, en raison de la présence de sulfures solubles, non détectés par la sonde. Leur concentration n'est pas parallèle à celle des matières en suspension, ce qui engendre sporadiquement un noircissement de l'effluent traité.
Il faut accroître la consommation de chlorosulfate ferrique pour que cette coloration noire n'apparaisse plus, mais la dose forfaitaire de réactif rend alors le procédé prohibitif, tandis que l'eau traitée vire au rouille.

4° Installation fonctionnant au chlorosulfate ferrique et utilisant le sulfate d'alumine à titre de dépannage

Le passage temporaire au sulfate d'alumine d'une station fonctionnant initialement au chlorosulfate ferrique, se traduit rapidement par un noircissement intense de l'effluent décanté.
L'installation de traitement des eaux et des boues reste inodore tant qu'il reste un volant de fer dans le circuit (boues de fond du décanteur, boues du silo).
La remise en service du poste d'injection de chlorosulfate ferrique entraîne une amélioration de la qualité des eaux traitées, suivie peu après d'une désodorisation du silo à boues.

Les diverses combinaisons de réactifs qui ont été testées ont fait ressortir la supériorité de la triple association sel ferreux, air, sel ferrique. L'air s'avère nécessaire pour conférer à l'eau décantée un niveau de traitement élevé. Vis-à-vis du fer ferreux injecté conjointement, cet air agit comme oxydant au même titre que le chlore pour la préparation en usine du chlorosulfate ferrique. Économiquement, l'oxydation par l'air sur place est moins dispendieuse que celle par le chlore en usine. D'autre part, le sulfate ferreux qui contient 18.5% en poids de fer est plus économique de transport que le chlorosulfate ferrique (ou le chlorure ferrique), qui n'en contient que 12% (ou 14%).

Diverses combinaisons de réactifs ont été testées sur l'installation réelle d'épuration, afin de voir si les performances relatives du procédé ferroso-ferrique s'avéraient conservées. Les résultats pratiques de ces essais sont les suivants:

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La stabilisation par l'air des sulfures naissants dans le silo est impossible car l'azote entraîne par stripping des gaz sulfurés vers l'atmosphère. Aussi, certaines unités ont-elles dû recourir à l'oxygène pur pour maîtriser les odeurs.

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On a avantage parfois à injecter le sel ferreux en amont de l'usine d'épuration, dans l'égout lui-même, afin de désodoriser les eaux d'égout jusqu'à leur acheminement à l'usine d'épuration.

Il existe ainsi, pour chaque municipalité, une filière propre de traitement, qui n'est pas extrapolable à une autre municipalité. La spécificité de chaque filière repose sur l'harmoniation au niveau du projet, des paramètres essentiels suivants: longueur du réseau, taux de parasitage de l'égout par des eaux étrangères, présence ou non d'effluents industriels, notamment de rejets gras, teneur en ammoniaque de l'effluent (NH_4^+ forme avec Fe^{++} un sel assez stable vis-à-vis de certains oxydants), teneur en colloïdes protecteurs, etc.

La prise en compte de ces paramètres doit avoir pour objectif de rechercher les conditions économiques idéales de formation du complexe ferroso-ferrique.

Si le coût d'exploitation demeure inchangé par rapport à celui des autres solutions, le niveau de risques sanitaire ainsi que la souplesse et la sécurité d'exploitation s'en ressentent favorablement.

TOXIC ORGANICS IN WASTEWATER - OCCURRENCE, ANALYSIS AND TREATMENT

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FILTRABLE SOLIDS AS A DESIGN PARAMETER FOR WASTEWATER TREATMENT PROCESSES

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Currently wastewater treatment facilities are designed on the basis of SS (suspended solids) and BOD (biochemical oxygen demand) measurements. Unfortunately, both SS and BOD are nonspecific measures of the suspended and biodegradable constituents present in wastewater. If significant advances are to be made in the design of treatment facilities a more fundamental characterization of wastewater will be necessary. Because a significant fraction of the organic material present in wastewater is associated with suspended and colloidal solids, it is worthwhile to focus on properties of the particulate matter. It is the purpose of this paper to suggest that the use of the mass distribution of filtrable solids is an appropriate and practical measure to provide a more useful characterization of wastewater for the design of treatment processes.

BACKGROUND

The importance of particle size in treatment operations is well established in air pollution control work and in water treatment (3,7). It is, therefore, surprising that this parameter has not been used more extensively in the wastewater treatment field. In a classic study reported 25 years ago, Balmat studied the treatability of the particles in wastewater in selected particle size ranges (1). The results of his studies are reported in Table 1. As shown in Table 1, the value of the BOD rate constant increases as the particle size range decreases. The overall rate is dependent on the distribution of particle sizes in the wastewater. Based on these data, Balmat concluded that ". . this indicates that comminution of some wastes to a particle size of 1 micron or less may greatly increase their rate of biochemical decomposition, thereby permitting a higher loading at the waste treatment plant."

Table 1

BOD RATE CONSTANT VALUES FOR SELECTED PARTICLE SIZE RANGES IN WASTEWATER[a]

Fraction	Size range, microns	K rate,[b] d^{-1}
Settleable	> 100	0.08
Supracolloidal	1-100	0.09
Colloidal	0.08-1	0.22
Soluble	< 0.08	0.39
Overall rate constant		0.22

[a] Adapted from Reference 1.

[b] BOD rate constant (base 10)

Balmat's objective of producing a waste with particle sizes equal to or less than 1 micron can also be achieved by removing the larger particle sizes. Particles sizes greater than 1 micron could be removed by chemical additions to the incoming wastewater or by mechanical filtration using the pulsed-bed filter (4). A treatment process employing a pulsed-bed filter followed by a tower trickling filter will be used as a case study to illustrate the significance of Balmat's observations. However, before discussing the case study, it is appropriate to consider the measurement of particle sizes and to present some data on the distribution of filtrable solids in water and wastewater.

MEASUREMENT OF PARTICLE SIZES

Techniques for measuring the particle sizes in water and wastewater can be grouped into the following general categories 1) electronic; 2) light scattering; 3) microscopic; 4) flow fractionation; 5) sedimentation; and 6) filtration. Electronic particle counters operate by measuring the difference between the conductivity of the particle and solution. Particles are measured in terms of equivalent spherical diameters and the smallest particle size measurable is about 0.3 microns. Light scattering devices measure the equivalent spherical diameter of particles based on a projected cross-sectional area. The minimum measureable size is on the order of one micron. A flow fractionator (2) in which particles are separated according to size in a shallow rectangular channel can also be used to study the size distribution of particles in wastewater. With a light microscope the lower limit of resolution is about one micron. With a scanning electron microscope (SEM) particles as small as 0.05 microns can be studied. The particles in wastewater can also be separated by gravity or by centrifugation. This was the technique used by Balmat to obtain the size ranges reported in Table 1. Filtration using a series of membrane filters with various pore sizes can also be used to determine the sizes of particles in wastewater. Although all of the above methods are under study, particle sizes and distributions obtained using the SEM and filtration techniques will be used in this paper.

EXPERIMENTAL PROCEDURES

The particle size distributions in various water and wastewater samples were quantified using the filtration technique described below. The particles

were studied further using SEM micrographs. Each of these techniques is documented in the following discussion.

Mass Distribution of Filtrable Solids

The mass distribution of filtrable solids was determined using Nuclepore polycarbonate membrane filters. The following series of filter pore sizes was used: 12, 8, 5, 3, 1, and 0.1 micron. The 0.1 micron filter was selected for the lower limit because filters with smaller pore sizes are extremely difficult to work with analytically. Where it was suspected that significant amounts of smaller particles may be present, filters with pore sizes smaller than 0.1 micron were used.

Samples were gently agitated and aliquots were applied to preweighed filters. Triplicates were run for each data point and the sample volumes used were standardized to minimize clogging. In the case of treatment plant influent, the filtrate from the 12 micron filter was applied to the remaining filters. In most of the other cases, the samples were not prefiltered. Samples were dried in a 100°C oven for a minimum of 1 hour and cooled in a dessicator for at least 45 minutes and weighed. The concentration of solids larger than a given pore size was determined. By comparison of successive filters, the distribution of solids was determined.

Observation of Particles with SEM

Scanning electron microscopy of the various samples was prepared using Nuclepore polycarbonate membrane filters. Samples were filtered in series through a 5 and a 0.1 micron filter. The filter was placed in 2.5 percent glutaraldehyde in 0.1M phosphate buffer for at least 1 hour, dehydrated in acetone and dried in a critical point dryer. The samples were mounted on stainless steel stubs and sputter-coated with a 500 angstrom layer of gold. Sample evaluation was performed using an ISI-DS130 Scanning Electron Microscope. For size distribution analysis, a Zeiss Videoplan Image Analyzing Computer was used.

CHARACTERIZATION OF SOLIDS IN WATER AND WASTEWATER

To illustrate the characteristics of the particles in water and wastewater, samples from four surface waters and five wastewater treatment plants have been analyzed.

Filtrable Solids Distributions

Typical data on the particle sizes found in the water and wastewater samples are reported in Table 2. The percentage distribution of filtrable solids by size range, reported in Table 2, is based on the mass of solids retained on a 0.1 micron filter. Percentages are used so that comparisons can be made between samples.

In the river and lake samples both the total mass and distribution of filtrable solids will vary depending on the time of year. In the untreated wastewater samples, these parameters will vary with the time of day. For wastewater, as shown in Table 2, the mass distribution of solids is bimodal. The effect of wastewater treatment processes on the distribution of filtrable solids is considered later in this paper. It is interesting to note that in untreated wastewater from 15 to 30 mg/L of filtrable solids will be found in the size range between 0.1 and 1.0 micron. In most cases, a large fraction of these solids is not measured in the conventional suspended solids test because the pore size of Whatman GFC filter used in the test is about 1.2 microns (6). In some cases filtrable solids smaller than 0.1 micron were measured. The wastewater from Los Banos, for example, contains a high proportion of colloidal milk wastes. An additional 32 mg/L of filtrable solids were measured in the size range between 0.015 and 0.1 micron. These solids were not included in computations of the percentage distribution of filtrable solids.

Micrographs of Solids

Micrographs of the particles found in water and wastewater are presented in Figures 1 through 3. The values for the magnification and scale length in microns (represented by a bar length) are given in the captions below each micrograph as the second and third entries respectively. Surface water samples are shown in Figure 1. In the lake and river waters examined, the material found was primarily inorganic with some freshwater microorganisms present. Primary effluent samples from two wastewater treatment plants are shown in Figure 2. The solids have been separated into two fractions. Clearly, it is impossible to describe the filtrable solids by an equivalent spherical diameter due to the amorphous nature of the material. The use of filtrable solids distribution provides a more reasonable approach because it provides a

Table 2

TYPICAL DATA ON THE DISTRIBUTION OF FILTRABLE SUSPENDED SOLIDS IN VARIOUS WATER AND WASTEWATER SAMPLES

Sample (Date, time)	Conc,[a] mg/L	Percent of mass retained in size range					
		>0.1 <1.0	>1.0 <3.0	>3.0 <5.0	>5.0 <8.0	>8.0 <12.0	>12.0
Surface water[b]							
Sacramento River (8/9/83, 4 pm)	56.3	36.8	1.95	0	19.7	0	41.6
American River (8/15/83, 2 pm)	8.8	28.4	47.7	0	4.0	11.4	8.5
Lake Berryessa (8/15/83, 11 am)	13.5	51.8	21.5	0	3.7	23.0	0
Folsom Lake (8/15/83, 10 am)	3.5	50.0	0	0	28.6	21.4	0
Untreated Wastewater[c]							
UCD (7/14/82) 11 am)	62.2	12.5	12.9	5.8	3.8	6.1	58.8
UCD (7/21/82, 4 pm)	129.9	16.1	25.1	0	0	0	58.8
Las Vegas (8/3/83, 8:30 pm)	146.1	14.2	32.4	6.9	0	6.5	40.0
Las Vegas (8/3/83, 2 pm)	284.0	1.8	32.6	11.5	11.1	1.8	41.2
Los Banos (8/8/83, 2 pm)	268.0	20.5	18.7	6.7	3.0	10.1	41.0
Secondary Effluent							
UCD (7/22/82)	38.9	10.3	4.9	3.1	0	0	81.7
Las Vegas (8/3/83, 11 am)	25.5	29.4	39.2	31.3[d]			
Las Vegas (3/3/83, 2:30 pm)	31.8	0.9	34.6	64.4[d]			
Las Vegas (8/3/83, 9 am)	29.0	0.7	23.1	76.2[d]			

[a] Based on the total mass of suspended solids retained on an 0.1 micron filter.

[b] Surface water samples collected in the vicinity of Sacramento, CA.

[c] Untreated wastewater samples collected at University of California, Davis, Las Vegas, NV, and Los Banos, CA.

[d] Size range > 3.0

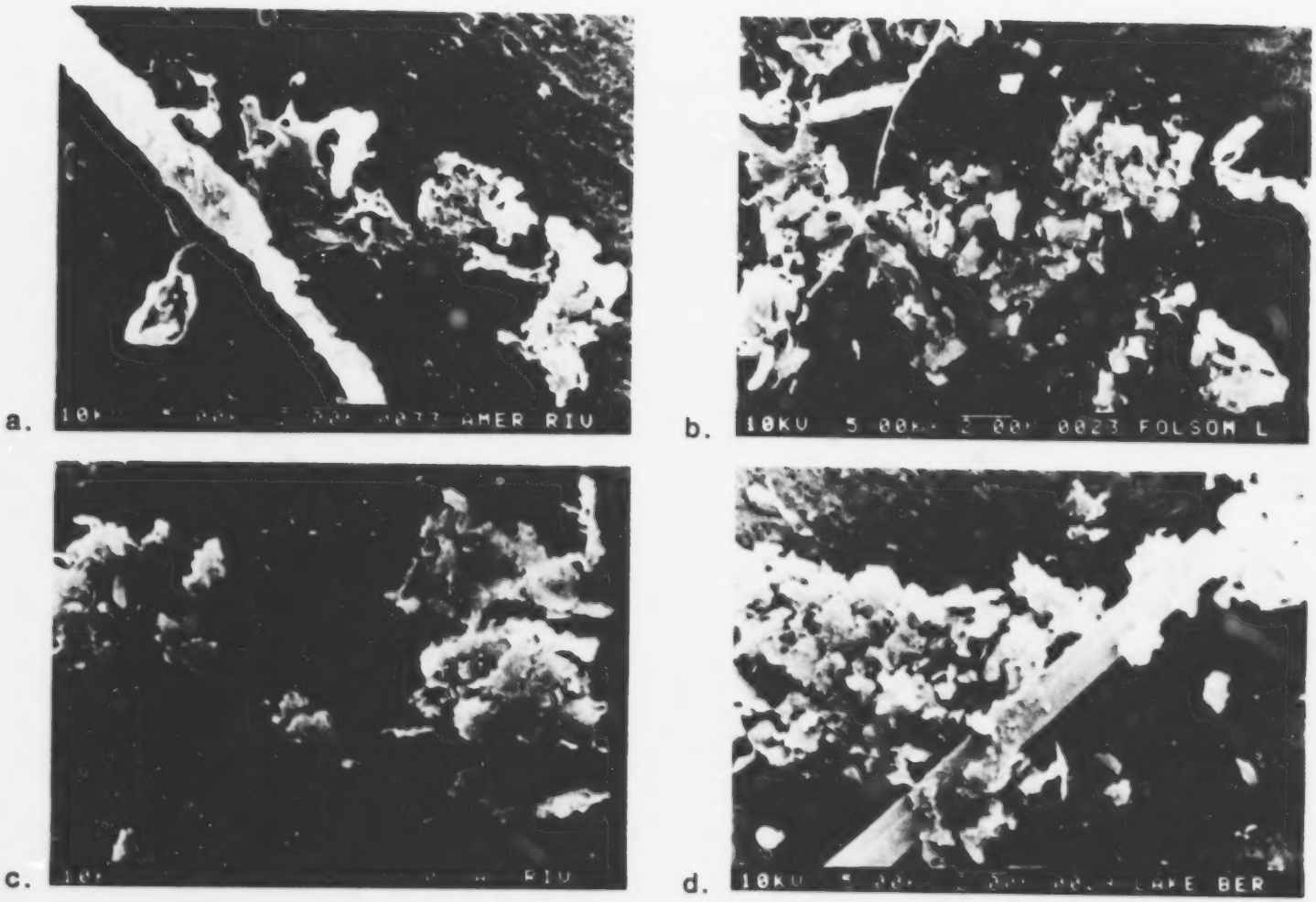

Fig. 1. Micrographs of filtrable solids in surface waters retained on a 0.1 micron membrane filter. a. American River, b. Folsom Lake, c. Sacramento River, d. Lake Berryessa. The length of the bar in figure captions represents 2 microns.

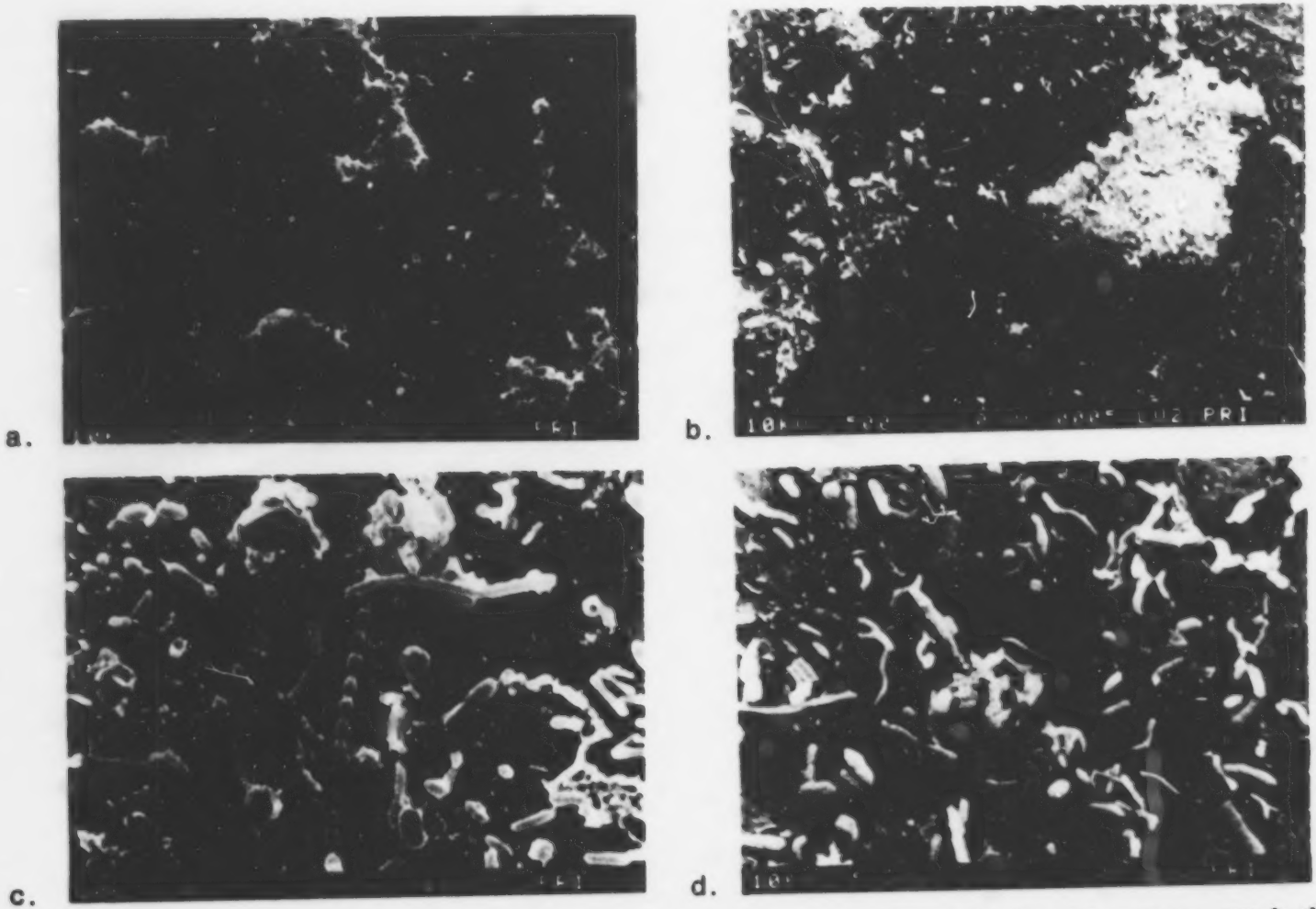

Fig. 2. Micrographs of filtrable solids in settled wastewater from the UC Davis (a,c) and Las Vegas, NV (b,d) treatment plants. a. Solids retained on 5.0 micron membrane filter, b. Solids retained on 1.0 micron membrane filter, c.,d. Solids retained on 0.1 micron membrane filter prefiltered through 5.0 micron membrane filter.

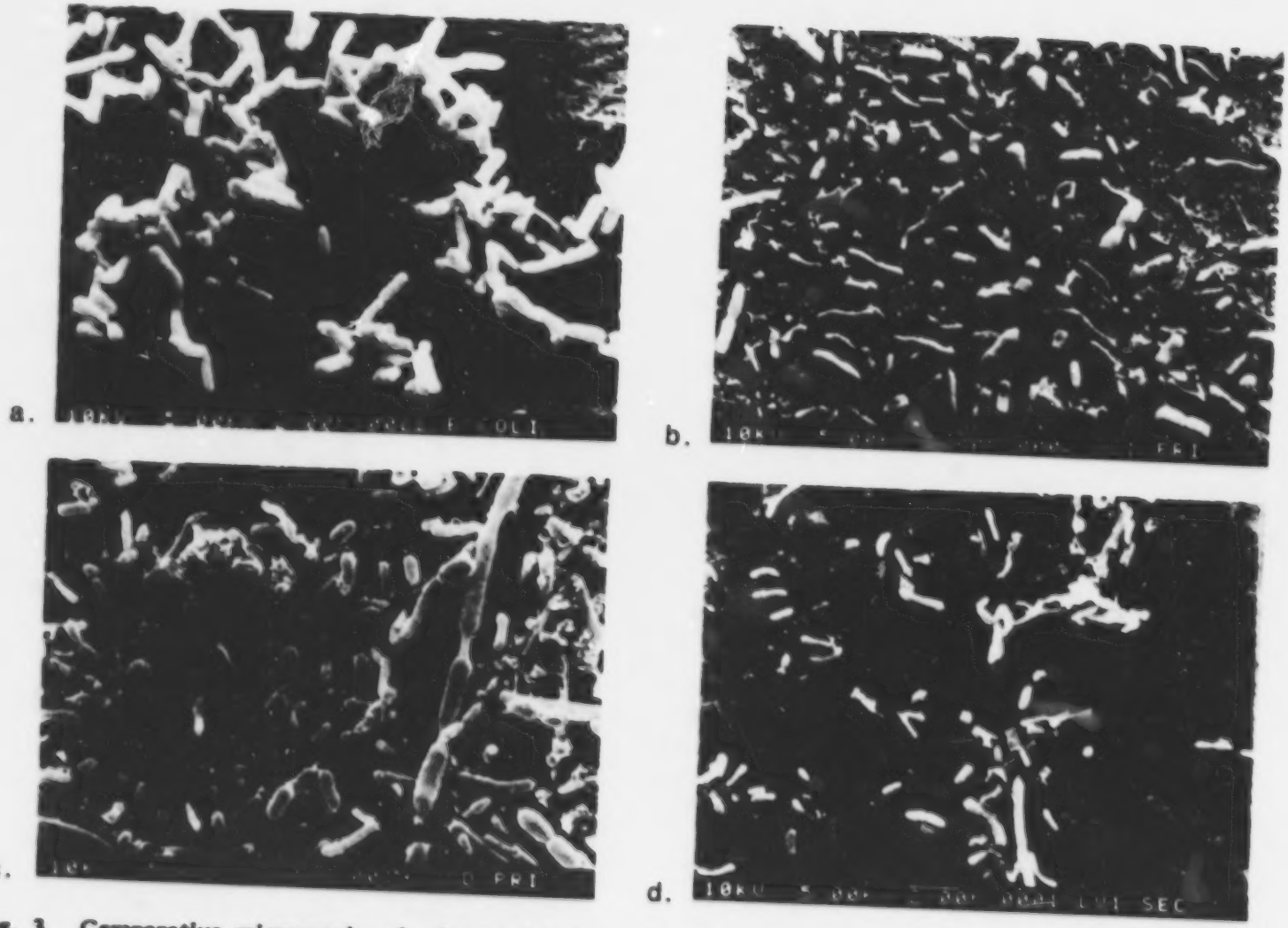

Fig. 3. Comparative micrographs of microorganisms. a. Pure culture of E coli, b. Primary effluent from Orange County, CA. c. Primary effluent from UC Davis treatment plant prefiltered through a 5.0 micron membrane filter, d. Secondary effluent from Las Vegas, NV.

reproducible measurement of the behavior of the solids in suspension. The results from electronic counters are variable depending on the conformation of the particle at the time it passes the detector.

Based on results from other studies, the majority of the filtrable solids in primary effluent are organic in nature. The treatability of the organic material is directly related to the size of the solids. The material retained on a 5 micron filter is difficult to treat biologically because it must initially be solubilized. Typically, the BOD to SS ratio for the larger solids is about 1.0. Solids smaller than 5 microns are typically microbial in nature. For purposes of comparison a pure culture of E coli is shown in Figure 3 along with typical micrographs of the organisms present in primary and secondary effluent. The BOD to SS ratio has the smaller particle ranges from 0.8 to 1.25 with a value of 1.0 being a reasonable approximation.

TREATMENT OF AN ALTERED WASTE: A CASE STUDY

In the previous section, data were presented on the distribution of particle sizes in water and wastewater. It is the purpose of this section to illustrate the benefits in treatment that result when the distribution of filtrable solids in water is altered before biological treatment.

Treatment Process Flowsheet

The treatment process flowsheet to be discussed is illustrated in Figure 4. As shown, a pulsed-bed filter is used to filter primary effluent before it is applied to a tower trickling filter. The operation of the pulsed-bed filter has been discussed elsewhere (4) and will not be repeated in this paper. What is of importance is what happens to the distribution of filtrable solids in the wastewater after filtration and how the changes that occur affect the biological treatment process.

Five tower filters, packed with a plastic medium, were operated with filtered primary effluent at hydraulic application rates of 40, 60, 80, 120, and 160 $m^3/m^2 \cdot d$ (note $m^3/m^2 \cdot d \times 24.2 = gal/ft^2 \cdot d$) without recycle. Effluent samples were collected and analyzed for SS and BOD without settling. For purposes of comparison, another tower trickling filter was used to treat unfiltered primary effluent at a filtration rate of 80 $m^3/m^2 \cdot d$, also without recycle.

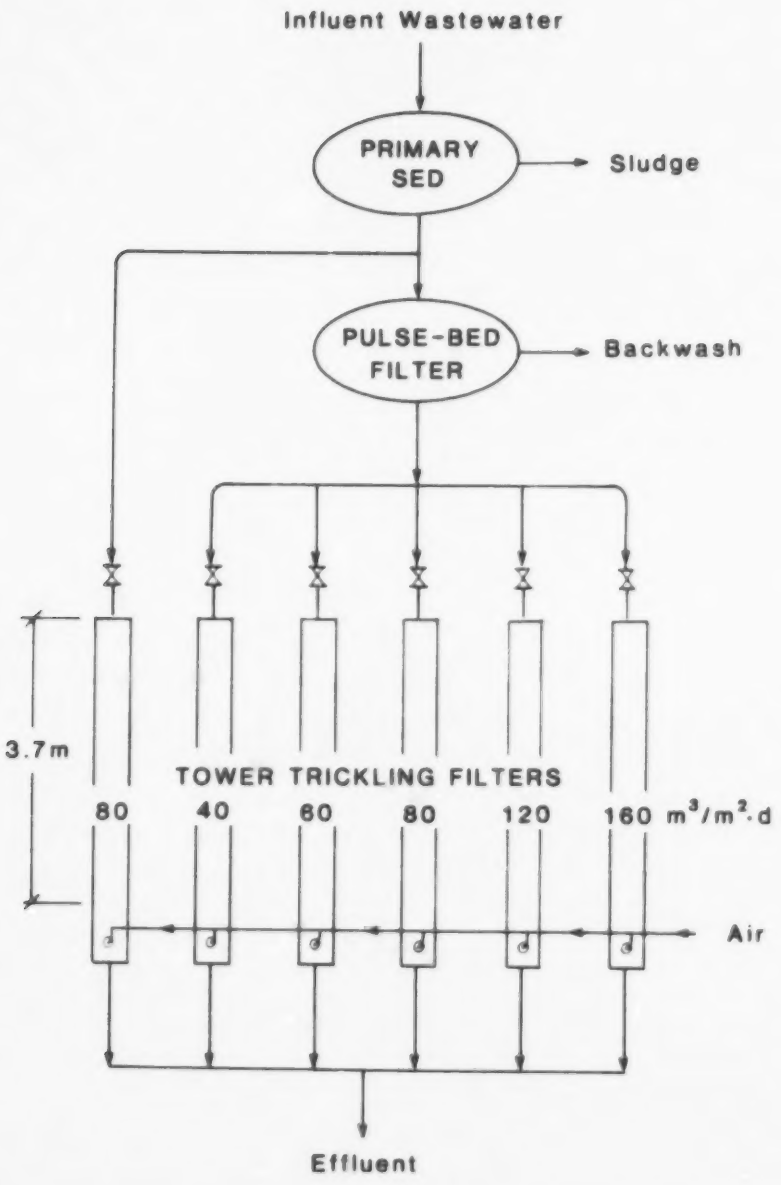

Fig. 4. Flowsheet for pilot plant employing pulsed-bed filter followed by tower trickling filters.

Treatment Performance

Performance data based on the nonspecific parameters of SS and BOD are presented along with data on the distribution of filtrable solids.

SS and BOD Removal Data Two pulsed-bed filters have been in continuous operation for over four years at the U.C. Davis campus wastewater treatment plant. Over this time period, the average removal of suspended solids, using a 0.65 mm sand as the filtering medium and a filtration rate of 80 $L/m^2 \cdot$ min, has been about 60 percent. Starting with a primary effluent of about 90 mg/L, the suspended solids in the filtered primary effluent have averaged about 32 mg/L. The corresponding average values for BOD are 140 and 70 to 80 mg/L in the filter influent and effluent respectively.

The effluent BOD from the five tower trickling filters receiving filtered primary effluent averaged about 30 mg/L regardless of the application rate. A representative set of operating data for the tower filters is presented in Table 3. Also included in Table 3 are influent and effluent data on suspended solids and filtered BOD.

Filtrable Solids Distribution Data on the distribution of particle sizes in the influent wastewater and the effluent from the primary sedimentation tank, pulsed-bed filter, and tower trickling filters are presented in Table 4. As noted in the discussion of Table 2, the distribution of filtrable solids in Table 4 is based on the mass of solids retained on the 0.1 micron filter. An approximate estimate of the corresponding suspended solids as measured in the conventional test (6) can be made by summing the mass of solids greater than 1 micron.

The effect of the pulsed-bed filter in altering the particle size distribution of the suspended solids applied to the tower trickling filters can be assessed comparing the distributions for the effluent from the primary sedimentation tank and the pulsed-bed filter. What has occurred is that the larger material in the primary effluent has been removed by the filter. Now, instead of having a bimodal particle size distribution, the particle size distribution in the filter effluent is skewed to the left, with the mode being somewhere around 1 micron.

Analysis of Process Performance

According to the findings presented by Balmat, the altered waste produced using the pulsed-bed filter should be quite treatable as the majority of the solids

Table 3

TYPICAL PERFORMANCE DATA FOR TOWER TRICKLING FILTERS TREATING FILTERED AND UNFILTERED PRIMARY EFFLUENT

Item	Concentration in mg/L at indicated hydraulic loading rate in $m^3/m^2 \cdot d$					
	40	60	80	80[a]	120	160
Influent						
SS	23	32	24	88	34	26
BOD, total	54	76	76	—	71	49
BOD, filtered	37	59	52	—	55	38
Effluent						
SS	17	22	30	80	36	28
BOD, total	20	32	30	—	31	25
BOD, filtered	12	19	15	—	13	13

[a] Unfiltered primary effluent

Table 4

TYPICAL DATA ON THE MASS DISTRIBUTION OF FILTRABLE SUSPENDED SOLIDS IN VARIOUS WASTEWATER SAMPLES[a]

Sample	Conc,[b] mg/L	Percent of mass retained in size range						
		> 0.1 < 1.0	> 1.0 < 3.0	> 3.0 < 5.0	> 5.0 < 8.0	> 8.0 < 12.0	> 12.0	> 35
Influent wastewater	220 (150-400)	10 (4-12)	8 (6-10)	2 (1-6)	1 (1-4)	3 (2-8)	16[c] (12-24)	60 (50-70)
Primary effluent	90 (60-120)	22 (10-30)	20 (16-25)	4 (2-12)	3 (2-8)	8 (4-16)	43 (30-60)	--
Filtered primary effluent	50 (25-60)	43 (30-55)	32 (20-40)	3 (0-10)	2 (0-8)	5 (2-12)	15 (10-20)	--
Trickling filter effluent	42 (30-50)	40 (35-55)	5 (2-10)	6 (2-10)	3 (0-5)	1 (0-2)	45 (30-55)	--

[a] All samples collected at the University of California, Davis wastewater treatment plant where the university pilot plant facilities are located.

[b] Based on the total mass of filtrable suspended solids retained on an 0.1 micron filter.

[c] Size range > 12, < 35

are in the 0.1 to 3.0 micron range. The performance of the filters can be assessed by comparing the loadings on the experimental trickling filters to those reported in the literature. Such a comparison is presented in Table 5. As shown, the hydraulic loading rates for the experimental filters would place them in the super-rate category. With the exception of the filter with the lowest hydraulic application rate, the organic loading rates to the other filters are all in the super-rate category. It is important to remember that such performance was achieved without any recycle, and that the effluent quality met secondary standards without settling.

The improved performance of these filters can be explained on the basis of particle size. By removing the larger particles that are more difficult to treat (see Table 4), the remaining particles are more easily processed. Particles in the size range from 0.1 to 1.0 micron are readily adsorbed. Solids in the 1 to 3 micron size range are also adsorbed. In addition, the secondary forms (e.g. protozoa) , which are an important part of the biological ecosystem within the filter, can directly ingest material smaller than three microns. On the other hand, larger solids tend to pass through the filter as illustrated schematically in Figure 5 and in the micrographs presented in Figure 6. As shown in Figure 6, the large particles in the primary effluent pass through the filter essentially unaltered. As would be expected, the number of particles in the smaller size ranges are reduced significantly as they pass through the filter. In conventional filters, high recirculation ratios are needed to break down these particles into smaller particles that can be removed by adsorption. In addition to hydraulic shearing some enzymatic hydrolysis also occurs in the bulk fluid as the particles are transported through the filter.

Another significant observation on the importance of altering the particle size distribution before biological treatment can be deduced by considering the two trickling filters shown schematically in Figure 6. As shown one filter is used to treat unfiltered primary and one filter is used to treat filtered primary. If the BOD removal is considered, it will be found that the value is about the same for both filters. This result is puzzling until it is realized that essentially the same particles are being removed in both filters. The difference is that the large particles are still present in the primary effluent whereas these particles are not present in the filtered primary effluent.

Table 5

COMPARISON OF LITERATURE LOADING VALUES FOR VARIOUS FILTERS TO LOADING VALUES FOR EXPERIMENTAL FILTERS[a]

Item	Hydraulic loading rate $m^3/m^2 \cdot d$	Organic loading rate $kg/m^3 \cdot d$
Low rate	1-4	0.08-0.32
Intermediate-rate	4-10	0.24-0.48
High-rate	10-40	0.32-1.0
Super-rate	40-200	0.80-6.0
Filter No.		
1	40	0.58[b]
2	60	1.23
3	80	1.64
4	120	2.30
5	160	2.12

[a] Literature values are from Reference 5

[b] 40 $m^3/m^2 \cdot d$ x 54 g/m^3 (see Table 3)/3.7 m (depth of filter)

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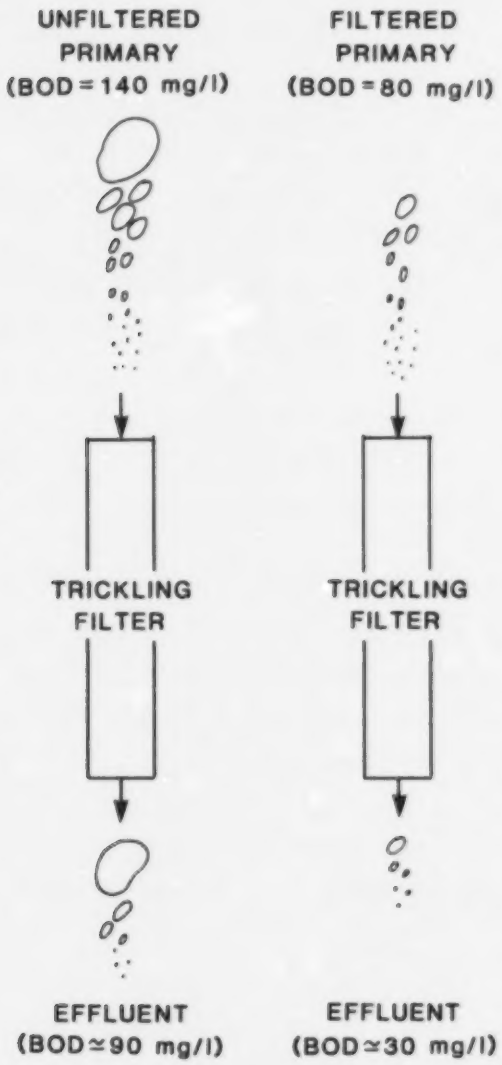

Fig. 5. Definition sketch for the treatment of unfiltered and filtered primary effluent with tower trickling filters.

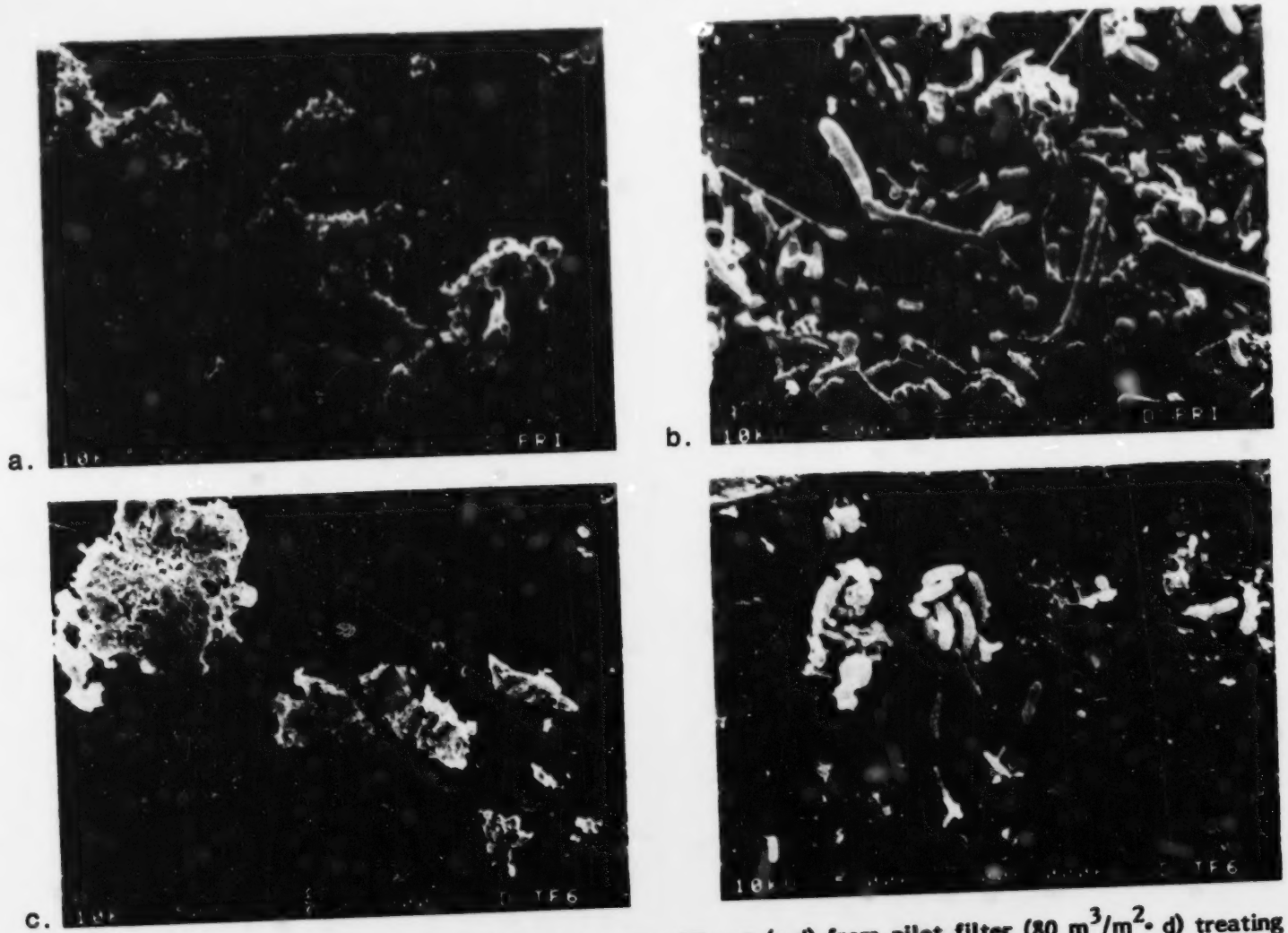

Fig. 6. Micrographs of trickling filter influent (a,b) and effluent (c,d) from pilot filter (80 $m^3/m^2 \cdot d$) treating unfiltered primary effluent at UC Davis treatment plant. a.,c. Solids retained on 5.0 micron membrane filter. b.,d. Solids retained on 0.1 micron membrane filter prefiltered through 5.0 micron membrane filter.

Because the large particles in the primary effluent pass through the trickling filter, they contribute to the effluent BOD. The presence of these large solids is what has given conventional trickling filters an unearned reputation of not being reliable with respect to effluent quality. It has been found that the fraction of large particles is proportional to the strength of the incoming wastewater. Thus as the strength of the incoming wastewater increases, more of these large particles pass through filter and the quality of the effluent deteriorates. If these particles were removed early in the treatment process, a high quality effluent could be produced even though the loading to the filter has increased significantly.

DISCUSSION

The benefits achieved in biological treatment by altering the particle size distribution of the suspended solids in primary effluent have been demonstrated using tower trickling filters. Similar results have been obtained from preliminary tests with the rotating biological disk and activated sludge process. The reported findings as discussed previously are consistent with the findings reported by Balmat and serve to underscore the importance of particle size.

Knowledge of how the solid material in the influent wastewater is distributed on the basis of size has important design implications. For example, if large solids dominate the distribution and there are essentially no particles in the size range between 0.1 and 1.0 microns it may be possible to achieve a high degree of treatment with chemical addition or with filtration. If the majority of the solids are in the size range between 0.1 and 1.0 microns enhanced settling or filtration offer limited benefits.

SUMMARY

The performance of biological treatment processes can be enhanced through the removal of suspended solids larger than 1.0 micron. To develop and design more effective treatment processes it is suggested that filtrable solids distribution measurements be made along with the measurement of SS and BOD. The use of membrane filters for this purpose is well documented and within the reach of every design engineer.

ACKNOWLEDGEMENTS

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ÉPURATION DES EAUX USÉES DOMESTIQUES PAR LAGUNAGE NATUREL

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INTRODUCTION

Le lagunage naturel est très employé dans de nombreux pays pour l'épuration des eaux usées domestiques, surtout pour des collectivités de quelques centaines à quelques milliers d'habitants. Son caractère rustique, son exploitation et son entretien faciles et économiques ne sont plus à démontrer.

Au Canada, et surtout au Québec, cette technique d'épuration des eaux est loin d'avoir conquis ses lettres de noblesse. Le rejet des eaux usées dans une fosse de dimensions empiriques et le manque d'intérêt de la part des autorités municipales, qui négligent souvent le peu d'entretien qu'un tel système exige, ont contribué à cause des faibles rendements obtenus à maintenir un doute sur la faculté du lagunage à épurer les eaux usées en région froide[1].

La section du Génie de l'Environnement de l'École Polytechnique, en collaboration avec la Société d'Énergie de la Baie James, a étudié le rendement et le comportement de deux étangs facultatifs non aérés à décharge contrôlée constitués chacun de deux bassins en série. L'étude s'est effectuée au Chantier La Grande 4 à la Baie James et s'est échelonnée sur trois étés consécutifs (1979 à 1981). Les résultats complets de cette étude sont compilés dans trois rapports (Héroux, 1979; Tridi, 1980; Leblanc, 1983a).

La première partie de ce document présente les principaux rendements obtenus durant cette étude. La seconde partie a pour objet de proposer des lignes directrices concernant les études préliminaires à réaliser, ainsi que des normes de conception pouvant s'appliquer à un projet de lagunage naturel en région froide.

ÉTUDE DES ÉTANGS DU CHANTIER LG-4

1.1 Description des installations

Le Chantier LG-4 se situe à 53° 54' de latitude Nord et à 73° 31' de longitude Ouest. Il se compose de deux parties: le village et le campement. Chacun d'eux possède un système de lagunage facultatif non aéré à décharge contrôlée composé de deux bassins en série. Les caractéristiques des deux étangs sont présentées au tableau 1 et leurs aspects aux figures 1 et 2.

---

1 région froide: région où les lagunes sont recouvertes de glace et de neige suffisamment longtemps pour justifier un emmagasinage des eaux usées durant l'hiver.

TABLEAU 1

CARACTÉRISTIQUES DES ÉTANGS DU CHANTIER LG-4

BASSIN	HAUTEUR (cm)		SUPERFICIE (ha)		VOLUME (m^3)	
	MINIMUM	MAXIMUM	MINIMUM	MAXIMUM	MINIMUM	MAXIMUM
Village A	1,25	2,70	2,40	2,63	288 800	65 400
Village B	2,25	4,95	2,00	2,60	45 000	106 500
Campement A	2,10	3,48	6,66	7,10	140 000	240 000
Campement B	1,80	3,18	2,80	3,58	50 400	95 000

	TEMPS DE RÉTENTION (mois)	DISTANCE DE LA ZONE D'HABITATION (m)
VILLAGE	11 à 14	1 000
CAMPEMENT	5,6 à 10	800

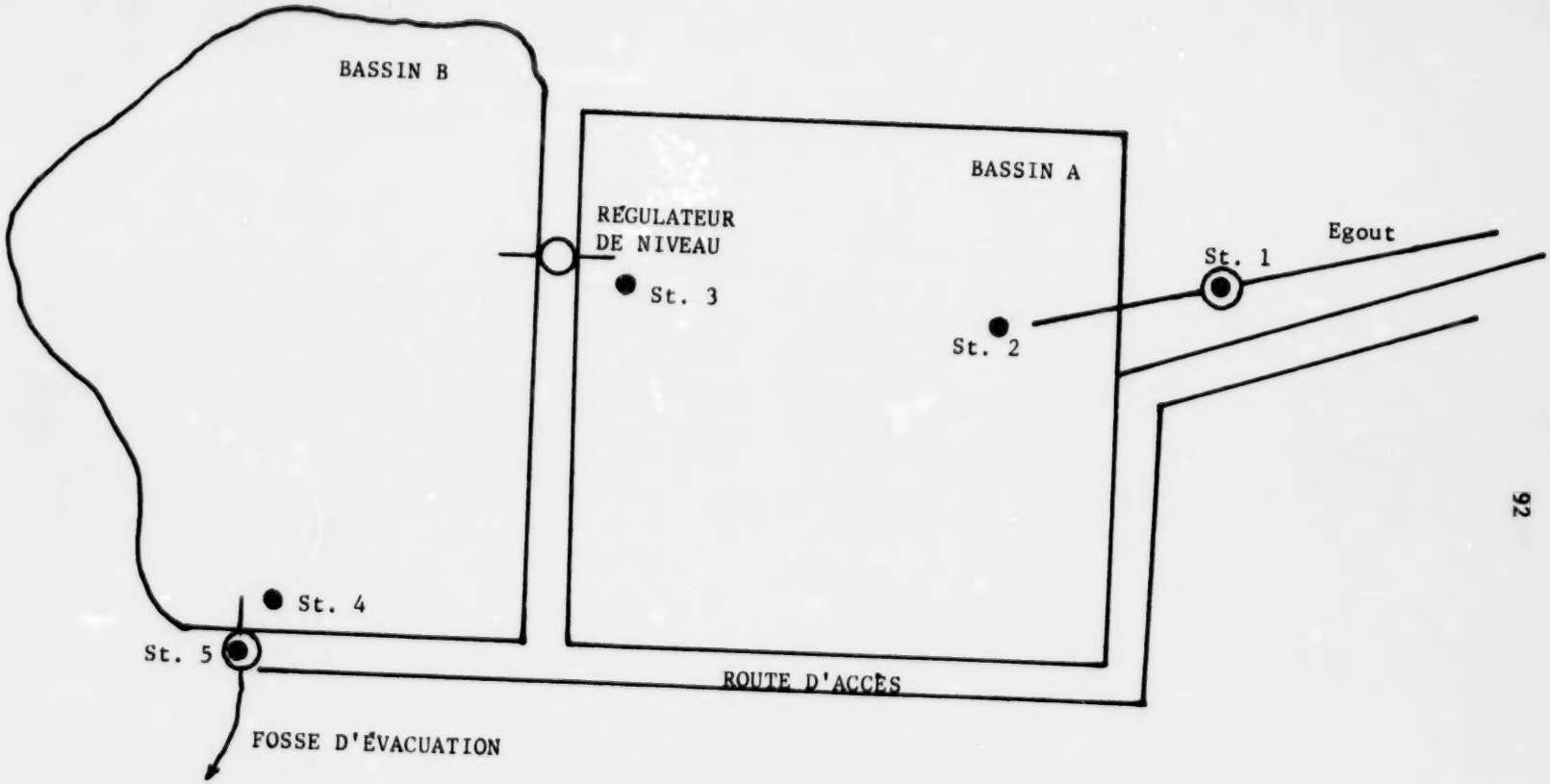

FIGURE 1. ETANG DU VILLAGE

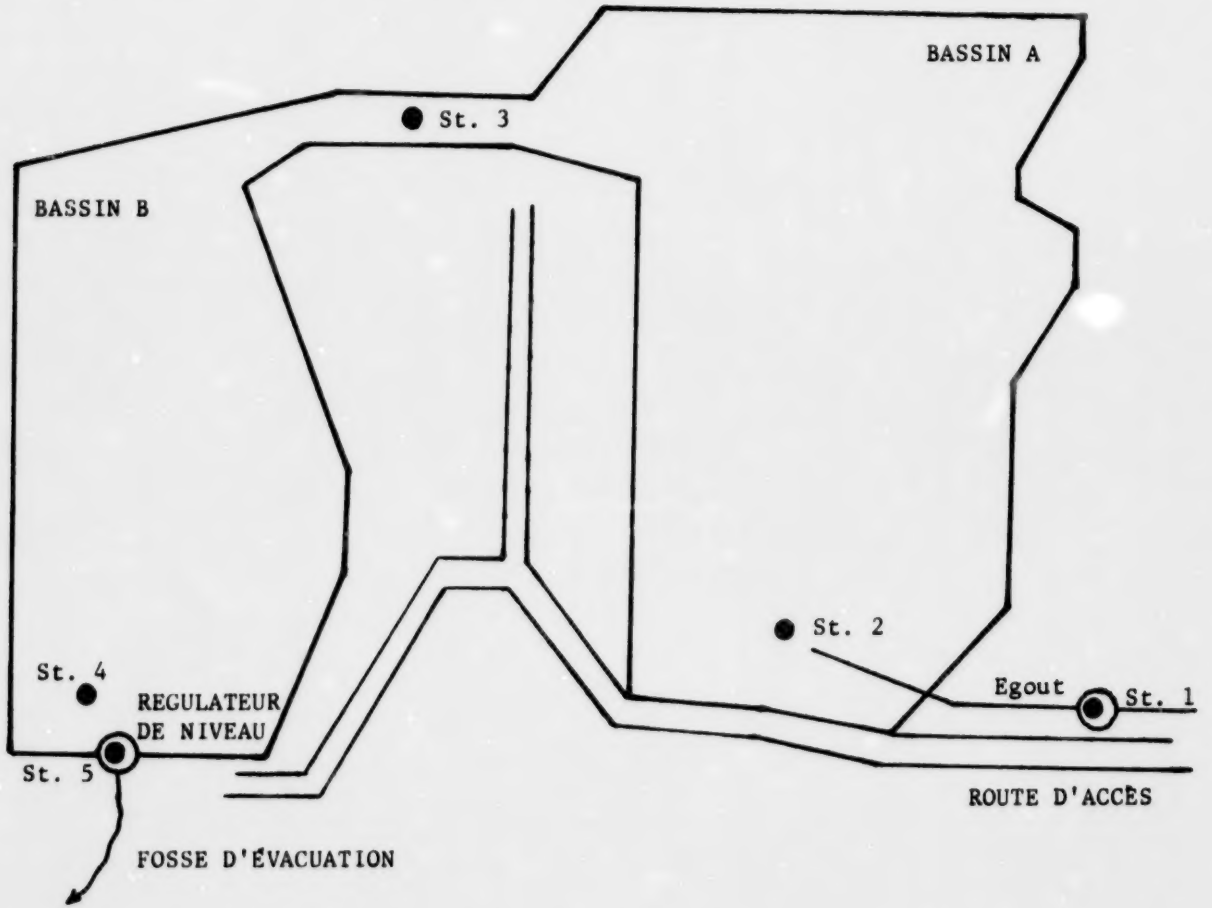

FIGURE 2. ÉTANG DU CAMPEMENT

1.2 Résultats et interprétations

L'étude menée à la Baie James comportait plusieurs volets dont les principaux portaient sur:

- la consommation de l'eau potable;
- la variation dans le temps et l'espace du pH, de l'oxygène dissous et de la température;
- la modélisation au niveau de la transformation des principaux nutriments;
- l'efficacité du lagunage à épurer les eaux, c'est-à-dire le rendement.

Nous limiterons cet exposé au volet concernant le rendement obtenu sur les principaux paramètres physico-chimiques.

Les deux installations d'épuration sont entrées en fonction durant la saison estivale de 1979. L'accumulation des eaux usées dans les bassins a débuté au printemps pour se poursuivre jusqu'à l'ouverture des vannes à l'effluent le 26 juillet à l'étang du campement et le 28 juillet à celui du village. Étant donné la mise en fonction tardive des étangs, il n'y a eu que très peu d'analyses physico-chimiques pour évaluer le rendement des étangs (tableau 2) durant la saison estivale de 1979.

En 1980, la campagne d'échantillonnage a été un peu plus élaborée. L'ouverture des vannes à l'effluent a eu lieu au début du mois de mai. La population du chantier s'est accrue, augmentant ainsi la charge. Le rendement obtenu sur les principaux paramètres physico-chimiques est présenté au tableau 3.

En 1981, période de pointe au chantier, la période d'échantillonnage s'est échelonnée sur 6 mois, ce qui a permis d'étudier les deux étangs sur un cycle complet, soit du dégel au gel des bassins, de l'ouverture à la fermeture des vannes à l'effluent. Les rendements présentés au tableau 4 sont des moyennes car, en réalité, ils varient beaucoup dans le temps et l'espace. En général, ils suivent l'évolution de la concentration des algues et de la température de l'eau (en moyenne, durant les mois de juillet et août 1981, la température de l'eau a été de 17,7 C). Le tableau 5 permet de constater cette variation de rendement dans le temps et l'espace. En fait, la dynamique de la population algale et bactérienne contrôle le système épurateur et donc le rendement.

Durant l'ensemble de l'étude, des odeurs n'ont été signalées qu'à de rares occasions (au dégel) et à des endroits bien précis (à l'entrée du premier bassin). Les habitants du chantier n'ont jamais été incommodés. Au point de vue esthétique, à part la coloration verte de l'eau et la remontée des boues à l'entrée du premier bassin lors du dégel, rien ne pouvait indiquer que les étangs étaient des systèmes d'épuration des eaux usées.

En général, le rendement obtenu durant les trois années de l'étude rencontre les principales normes recommandées en Amérique du Nord. On y constate que les performances sont très satisfaisantes surtout pour la DBO_5, la DCO, l'azote et le phosphore. En ce qui concerne la concentration des matières solides totales et des matières en suspension, elles sont surtout fonction de la capacité de sédimentation des algues du dernier bassin. En fait, le rendement d'une lagune, en été, est comparable à n'importe quel traitement secondaire conventionnel lorsque les normes de conception et de construction sont respectées. Au niveau de l'efficacité des étangs pour

TABLEAU 2

RENDEMENT DES ÉTANGS DU CHANTIER LG-4, ÉTÉ 1979

Période d'échantillonnage: 10 juillet au 29 août 1979

	CAMPEMENT	VILLAGE
Population (hab)* :	1 036	393
Consommation (ℓ/hab.d)*:	464	873
Charge (Kg DBO_5/ha.d) :	5,0	4,9

PARAMÈTRES	CAMPEMENT			VILLAGE		
	AFFLUENT	EFFLUENT	RENDEMENT (%)	AFFLUENT	EFFLUENT	RENDEMENT (%)
DBO_5 (mg/ℓ)	103,8 (2)**	5,6 (2)	94,6	68,5 (1)	1,5 (2)	97,8
DCO (mg/ℓ)	318,0 (5)	84,6 (5)	73,4	244,8 (5)	46,0 (4)	81,2
Matières solides totales (mg/ℓ)	475,5 (4)	166,5 (4)	65,0	362,0 (2)	109,5 (4)	69,8
Matières en suspension (mg/ℓ)	168,8 (4)	42,8 (4)	74,6	44,5 (2)	20,5 (4)	53,9
Azote total (mg/ℓ N)	26,8 (5)	3,2 (3)	88,1	39,1 (5)	3,2 (3)	91,8
Phosphore total (mg/ℓ P)	5,2 (5)	0,9 (5)	82,7	6,4 (4)	1,1 (5)	82,8

* moyenne hebdomadaire, janvier à août 1979 ( )** nombre d'échantillonnage

TABLEAU 3

RENDEMENT DES ÉTANGS DU CHANTIER LG-4, ÉTÉ 1980

Période d'échantillonnage: 13 mai au 7 août 1980

	CAMPEMENT	VILLAGE
Population (hab) :	1 887*	630**
Consommation (ℓ/hab.d) :	431*	373**
Charge (Kg DBO_5/ha.d) :	12,4	8,6

PARAMÈTRES	CAMPEMENT			VILLAGE		
	AFFLUENT	EFFLUENT	RENDEMENT (%)	AFFLUENT	EFFLUENT	RENDEMENT (%)
DBO_5 (mg/ℓ)	153,0 (5)***	31,5 (8)	79,4	176,3 (6)	13,5 (8)	92,3
DCO (mg/ℓ)	395,5 (5)	138,7 (6)	64,9	373,4 (5)	93,2 (6)	75,0
Matières en suspension (mg/ℓ)	204,4 (7)	57,7 (10)	71,8	101,0 (7)	30,0 (10)	70,3
Azote total (mg/ℓ N)	26,6 (7)	7,7 (10)	71,1	31,4 (8)	5,4 (10)	82,8
Phosphore total (mg/ℓ P)	8,8 (6)	3,0 (8)	65,9	11,4 (7)	2,3 (8)	79,8
Coliformes fécaux (colonies/100ml)	$1,7x10^7$ (7)	$1,5x10^3$ (8)	56,1 ****	$2,6x10^6$ (7)	115 (9)	67,9 ****

* moyenne hebdomadaire, mai à juillet 1980
** moyenne hebdomadaire, mai et juin 1980
( )*** nombre d'échantillonnage
**** rendement logarithmique

TABLEAU 4

RENDEMENT DES ÉTANGS DU CHANTIER LG-4, ÉTÉ 1981

Période d'échantillonnage: 17 mai au 28 octobre 1981

	CAMPEMENT	VILLAGE
Population (hab)* :	1 687	693
Consommation (ℓ/hab.d)*:	551	375
Charge (Kg DBO_5/ha.d) :	37,5	9,9

PARAMÈTRES	CAMPEMENT			VILLAGE		
	AFFLUENT	EFFLUENT	RENDEMENT (%)	AFFLUENT	EFFLUENT	RENDEMENT (%)
DBO_5 (mg/ℓ)	403,9 (16)**	27,6 (14)	93,2	182,1 (16)	15,8 (13)	91,3
DCO (mg/ℓ)	683,3 (16)	173,3 (15)	74,6	529,0 (16)	78,6 (14)	85,1
Matières solides totales (mg/ℓ)	646,8 (16)	254,2 (14)	60,7	690,8 (16)	190,4 (14)	72,4
Matières en suspension (mg/ℓ)	316,6 (16)	84,0 (15)	73,5	231,1 (16)	24,0 (14)	89,6
Azote Kjeldahl (mg/ℓ N)	35,1 (16)	13,2 (15)	62,4	56,3 (16)	6,2 (14)	89,0
Phosphore total (mg/ℓ P)	8,0 (16)	3,5 (15)	56,3	11,0 (16)	3,1 (14)	71,8
Coliformes fécaux (colonies/100ml)	$6,2x10^6$ (15)	$2,9x10^4$	34,3 ***	$8,5x10^3$ (16)	$1,2x10^3$ (14)	55,6 ***

* moyenne mensuelle, 1 novembre 1980 au 31 octobre 1981

*** rendement logarithmique

( )** nombre d'échantillonnage

TABLEAU 5

VARIATION DU RENDEMENT DANS LE TEMPS ET L'ESPACE

PARAMÈTRES	EFFLUENT	1er BASSIN		EFFLUENT	2ième BASSIN	
	*	**	***	*	**	***
VILLAGE						
Matières en suspension (mg/ℓ)	27,0	73,3	54,2	11,3	27,7	19,0
DCO (mg/ℓ)	97,4	173,0	138,0	70,4	85,3	67,1
Azote Kjeldahl (mg/ℓ N)	13,9	9,6	11,6	8,9	4,2	6,1
CAMPEMENT						
Matières en suspension (mg/ℓ)	42,6	89,7	129,0	40,2	98,2	109,0
DCO (mg/ℓ)	151,0	222,0	225,0	132,0	193,0	195,0
Azote Ammoniacal (mg/ℓ N)	7,2	4,8	3,3	6,7	3,8	3,5

* Période du 17 mai au 21 juin 1981
** Période du 5 juillet au 13 septembre 1981
*** Période du 20 septembre au 28 octobre 1981

l'enlèvement des coliformes, elle est bonne surtout à l'étang du village. Un troisième bassin aurait amélioré de beaucoup la qualité bactérienne de l'eau.

ÉTUDES PRÉLIMINAIRES [1]

2.1 Critères d'aptitude du site

L'analyse de photographies aériennes et une visite sur les lieux permettent de délimiter les zones susceptibles d'être aménagées.

On aura à l'esprit que:

- pour la protection de la population, l'étang doit être situé à au moins 0,5 km de la plus proche habitation;
- l'étang doit être situé dans un lieu qui offre une bonne protection contre les vents froids du nord et qui doit être orienté de préférence vers le sud pour recevoir davantage de lumière incidente et d'énergie solaire;
- les zones rocheuses susceptibles d'entraîner des risques de fuite et des plus values lors du terrassement doivent être rejetées.

Le milieu de rejet peut également influencer le choix du site. Un sol possédant un bon drainage, une rivière, un marais ou une tourbière sont autant de milieux pouvant recevoir l'effluent d'un étang et même compléter l'épuration.

2.2 Études approfondies du site

Une fois que le ou les sites potentiels sont délimités, il convient de pousser plus loin les études, surtout au niveau des sols, pour choisir l'emplacement idéal. Des études de géologie, de géotechnique et d'hydrogéologie permettront de préciser la structure du terrain en place et de caractériser les matériaux qui la composent. Les résultats de ces études permettront:

- d'arrêter définitivement le choix du site:
- d'examiner la possibilité de réemploi des déblais;
- de déterminer une norme de compactage pour le fond du bassin et pour les digues;
- de prévoir, s'il y a lieu, les techniques d'imperméabilisation;
- de ne pas compromettre l'exploitation actuelle ou future de la nappe souterraine;
- de connaître les conditions du pergélisol.

2.3 Caractéristiques des eaux usées

Avant d'entreprendre la conception des étangs, il convient de connaître la qualité et la quantité des eaux à traiter, car ces paramètres peuvent varier considérablement d'un endroit à un autre.

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1 Ce chapitre et les suivants sont un résumé d'un document préparé par Leblanc (1983b) sur le lagunage naturel et édité par le Centre d'Ingénierie Nordique de l'École Polytechnique.

Pour dimensionner convenablement un étang, il faut connaître la charge polluante moyenne (kg DBO_5/d) calculée sur 60 jours consécutifs lorsque la charge est à son plus haut niveau; on doit éviter de prendre la charge polluante maximale.

2.4 Études climatiques

Un relevé du climat de la région permet de connaître la date approximative de la fonte et de la prise des glaces, déterminant ainsi la durée de l'emmagasinage des eaux. Cette durée peut être évaluée également par l'observation des dates de fonte et de prise des glaces des petits lacs de la région considérée.

Des données relatives à la pluie, la neige, l'évaporation ainsi que l'infiltration permettront de calculer le bilan volumique lors de la conception.

Des données réparties sur plusieurs années s'avèrent nécessaires pour avoir une vue générale des conditions climatiques.

NORMES DE CONCEPTION

3.1 Ouvrages de pré-traitement

Les ouvrages de pré-traitement (dégrillage, dilacération, dessablage) ne sont pas indispensables au bon fonctionnement du lagunage (Agence de Bassin Loire-Bretagne, 1979). Ils doivent répondre à deux conditions:

- être suffisamment simples et rustiques pour ne pas entraîner de surcroît à l'investissement et être d'une exploitation et d'un entretien faciles;
- accepter les surcharges hydrauliques sans que leur efficacité soit trop compromise.

En fait, ces deux conditions forcent à limiter les ouvrages de pré-traitement.

3.1.1 Critières de conception. Plusieurs méthodes sont susceptibles d'être employées pour concevoir un étang facultatif. Parmi celles-ci l'on retrouve les équations empiriques et les modèles cinétiques. Cependant ces méthodes n'ont pas été développées en fonction d'un climat qui oblige l'emmagasinage des eaux usées durant une partie de l'hiver.

On préconise l'utilisation de deux paramètres permettant de dimensionner l'installation:

- la charge polluante (DBO_5);
- le volume d'eau à traiter.

D'après l'étude menée à la Baie James ainsi qu'une recherche bibliographique, la charge recommandée comme base de dimensionnement varie de 12 à 22 kg de DBO_5/ha x d selon qu'on se situe au nord ou au sud du 50ième parallèle. Ce critère de charge sera précisé dans la section suivante.

Au sujet du volume d'eau à traiter, ce paramètre est employé pour vérifier si la superficie de l'étang calculée à partir de la charge (en assumant une profondeur variant entre 2.0 et 2.5 m en hiver) donne un volume suffisant pour emmagasiner les eaux usées produites durant l'hiver.

3.2 Nombre, forme et superficie des bassins

Le nombre, la forme et la superficie des bassins dépendra du degré d'épuration que l'on veut obtenir. Ce dernier est fonction de la nature et

de l'utilisation du milieu récepteur, ce qui constitue d'ailleurs l'un des éléments fondamentaux d'une bonne politique de gestion.

Une loi générale dûment vérifiée est que le jeu de la concurrence vitale privilégie la flore bactérienne la mieux adaptée aux conditions du milieu au détriment des autres. Donc chacun des bassins se trouvera colonisé par une flore bactérienne spécifique des principaux stades de la transformation de la matière organique (Agence de Bassin Loire-Bretagne, 1979). Donc plus le degré d'épuration demandé est élevé, plus le nombre de bassins requis sera grand.

Dans tous les cas, le nombre minimum de bassins est de deux et idéalement de trois. Le premier bassin est considéré surtout comme un décanteur fonctionnant en aérobiose dans sa partie supérieure (2/3 du volume) et en anaérobiose dans sa partie inférieure. Pour maintenir ce caractère facultatif, la charge maximale sera de 30 kg DBO_5/ha x d. Si la superficie requise de bassin dépasse 5 hectares, il serait souhaitable de la répartir entre deux bassins fonctionnant en parallèle. La forme du/des bassin(s) devrait permettre une bonne diffusion de l'affluent et éviter les courts-circuits. Ringuelet (1982) propose un bassin de forme circulaire sans chicane dans lequel la réduction de la charge (kg DBO_5/ha x d) devrait être de l'ordre de 75%.

Les deux autres bassins en série ont un rôle de polissage et sont en aérobie totale. Ils auront une superficie et un temps de rétention identiques. Pour préserver ce caractère aérobie, la charge maximale sera de l'ordre de 5 kg DBO_5/ha x d si l'étang est situé au nord du 50ième parallèle et de 10 kg DBO_5/ha x d s'il est situé au sud du 50ième parallèle. Ce fractionnement de la charge entre le premier bassin et les autres revient à préciser le critère moyen de charge de 12 et celui de 22 kg DBO_5/ha x d. Au sujet de la forme des bassins, Ferrara et Harleman (1978) recommandent de favoriser un écoulement en piston dans les bassins secondaires et tertiaires; c'est-à-dire un rapport longueur/largeur minimum de 5 à 1 et idéal de 13 à 1. L'écoulement en piston pourra être obtenu en utilisant des digues d'homogénéisation (voir section 3.6).

Il est parfois possible de profiter du relief et d'utiliser une dépression ou même un étang déjà existant, ce qui permet de réduire les opérations de construction. L'étang ou la dépression devra quand même être divisé en trois parties indépendantes pour former les trois bassins.

Il reste que la conception générale sera basée sur l'utilisation la plus rationnelle des terrains disponibles dans le but de profiter au mieux des possibilités d'intégration au site.

3.3 Profondeur et gestion

La profondeur des bassins est assujetie à trois contraintes:

- éviter le gel jusqu'au fond des bassins;
- maintenir une zone aérobie importante dans le premier bassin et par conséquent dans les autres bassins;
- posséder la capacité d'emmagasiner les eaux usées durant la période d'hiver.

L'activité photosynthétique se situe entre 0 et 1 m sous la surface et est maximale dans le 30 cm supérieurs. La lumière pouvant atteindre 1.5 m, cette profondeur sera celle du premier bassin en période d'été. La profon-

deur normale d'opération des autres bassins variera entre 1.0 m et 1.25 m pour maintenir les conditions aérobies.

L'efficacité des étangs dépend en grande partie de la gestion des bassins au niveau des profondeurs de ces derniers. On préconise une vidange rapide au printemps des derniers bassins (c'est une eau déjà épurée en partie) pour atteindre le niveau normal d'opération le plus vite possible et ainsi favoriser les conditions qui permettront une forte croissance algale et bactérienne. Au cours de l'automne, une vidange graduelle des bassins permettra de diminuer la profondeur jusqu'à son niveau minimum au début du gel des étangs en vue de l'emmagasinage d'hiver. L'épaisseur de la couche d'eau devrait s'accroître au même rythme que l'accroissement de la couche de glace, ce qui permettrait d'éviter le gel des structures de sortie et du fond des bassins. Il est évident que cette gestion des bassins dépend de la capacité du milieu récepteur à répondre à ce scénario.

Le temps nécessaire à la stabilisation de la matière organique est d'environ 60 jours en été (Agence de Bassin Loire-Bretagne, 1979), ce qui traduit en fait la condition de durée nécessaire à la prise en compte de la charge organique par les bactéries en suspension. On devra donc vérifier si le système remplit cette condition.

3.4 Alimentation du premier bassin

Il subsiste une controverse entre les deux techniques les plus employées pour alimenter le premier bassin, soit:

- l'alimentation en surface;
- l'alimentation en profondeur.

L'alimentation en profondeur est préconisée par les auteurs, car d'après l'expérience acquise à la Baie James, un affluent en surface empêche la sédimentation des particules grossières (comme des immondices et papiers hygiéniques) entraînant une diminution de la photosynthèse et rendant le bassin complètement inesthétique. De plus, dans un étang où les eaux sont froides, l'alimentation en surface encourage un courant préférentiel plus chaud à la surface. Pour répartir la charge organique et favoriser un régime hydraulique qui empêche ou du moins réduit les zones d'eaux mortes, deux à trois conduites ancrées dans le fond du bassin et débouchant près d'un piège à boue sont recommandées (figure 3). Ces pièges sont indispensables pour ne pas réduire le volume utile du bassin à cause de la lenteur de la stabilisation des boues en milieu nordique (Labonté et Béron, 1976).

3.5 Communication entre bassins et structure de sortie

L'ouvrage de communication entre bassins et celui de structure de sortie qui semble le plus adapté aux conditions climatiques est le moine de pisciculture incorporé dans la digue (régulateur de niveau) (figure 4). Un moine permet de construire des bassins avec des niveaux de plans d'eau différents et de les régler à volonté, ce qui est de première importance pour l'emmagasinage d'hiver. Voici quelques recommandations pour la construction du moine:

- La conduite du côté amont doit être située au point le plus bas possible du bassin avec une prise d'eau entre 20 et 30 cm du fond. On parvient ainsi à réduire l'exportation des matières en suspension et en particulier du phytoplancton qui se trouve en densité maximale dans la tranche d'eau supérieure.

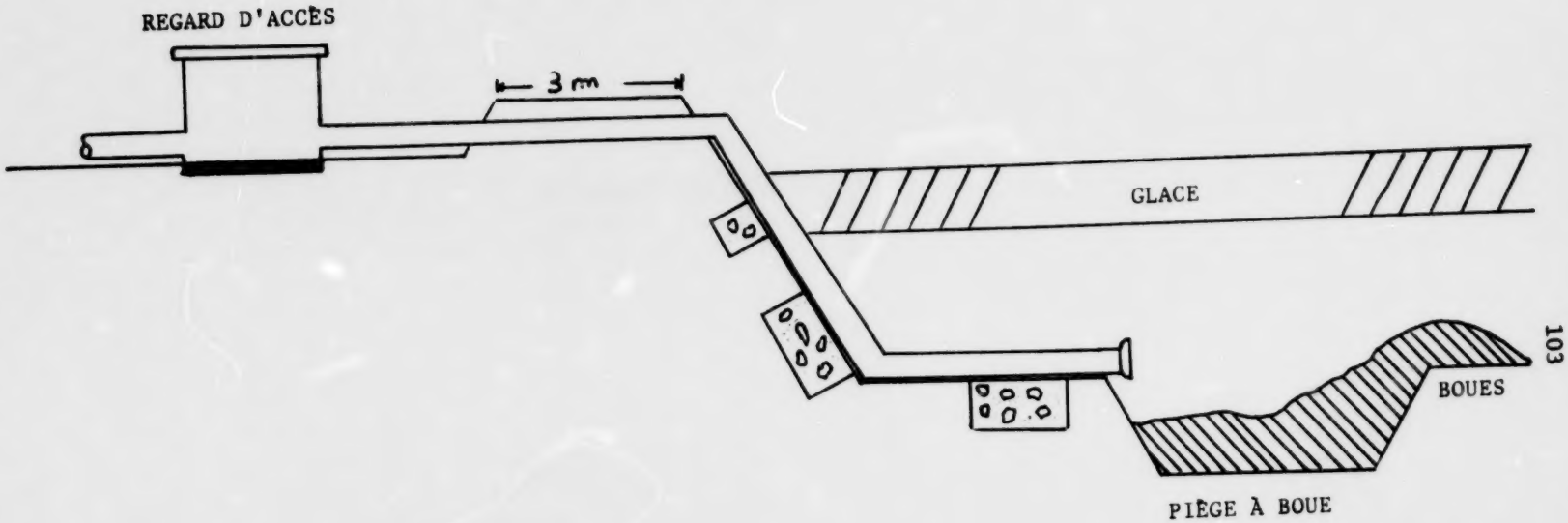


FIGURE 3. ALIMENTATION DU PREMIER BASSIN PAR LE FOND

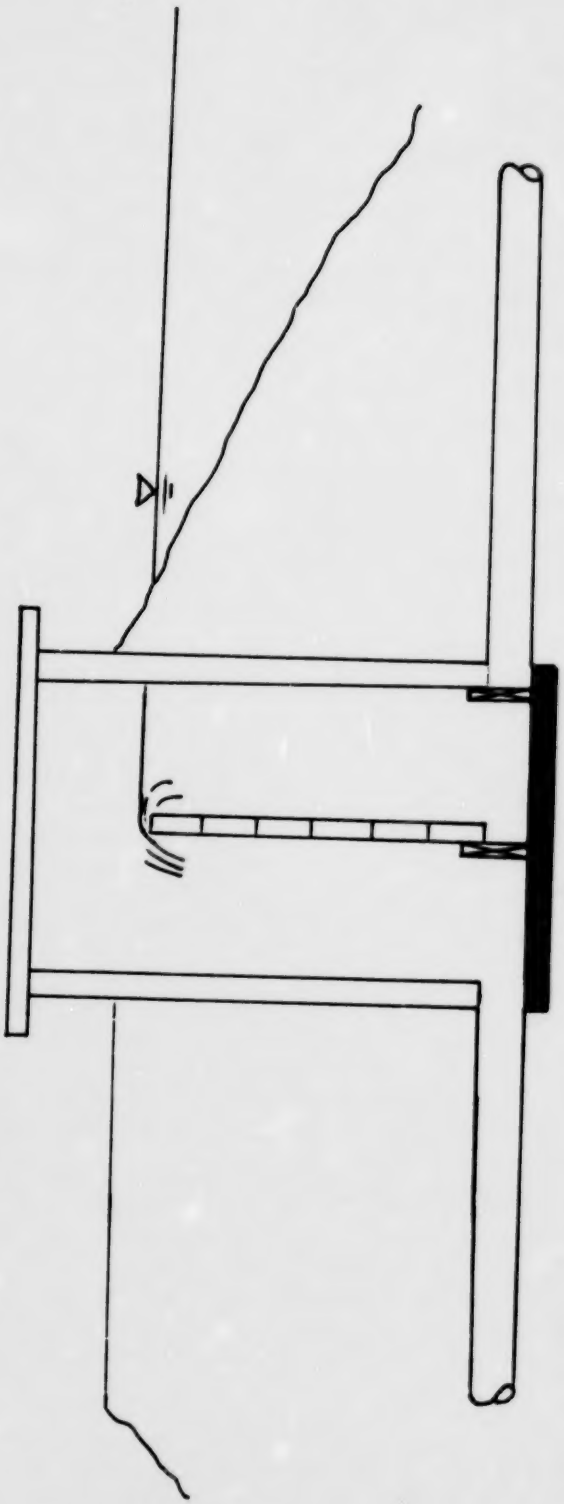


FIGURE 4. MOINE DE PISCICULTURE
(Régulateur de niveau)

- Du côté aval, le déversement à la surface du milieu récepteur est préférable. Entre bassins, ce déversement doit s'effectuer en profondeur à cause de la formation d'un bouchon de glace qui empêcherait le passage de l'eau entre les bassins durant l'hiver.
- L'utilisation de planches en bois dur adaptées après gonflement à l'humidité aux dimensions de l'ouvrage est nécessaire pour éviter les fuites.
- Une seule cloison de planches empilées est suffisante pour que l'ouvrage serve à la fois de trop plein et de vidange.

3.6 Digues d'homogénéisation

Lorsque les bassins secondaires et tertiaires ont des surfaces importantes, il est désirable de disposer en chicanes des digues rustiques d'homogénéisation pour favoriser l'écoulement en piston et limiter l'établissement de zones d'eaux mortes. On ne recherchera ni leur étanchéité, ni leur résistance à la pression des fluides; par contre elles devront résister à la glace. Selon Guerrin (1981), on les retrouve principalement sous quatre formes:

1- en terre: mais elles compliquent les terrassements et augmentent l'emprise foncière;
2- en éléments préfabriqués en béton maintenus par des poteaux rainurés;
3- en murs d'agglomérés bâtis sur semelle de béton;
4- en pieux fichés dans le sol et reliés par des palplanches.

3.7 Milieu récepteur

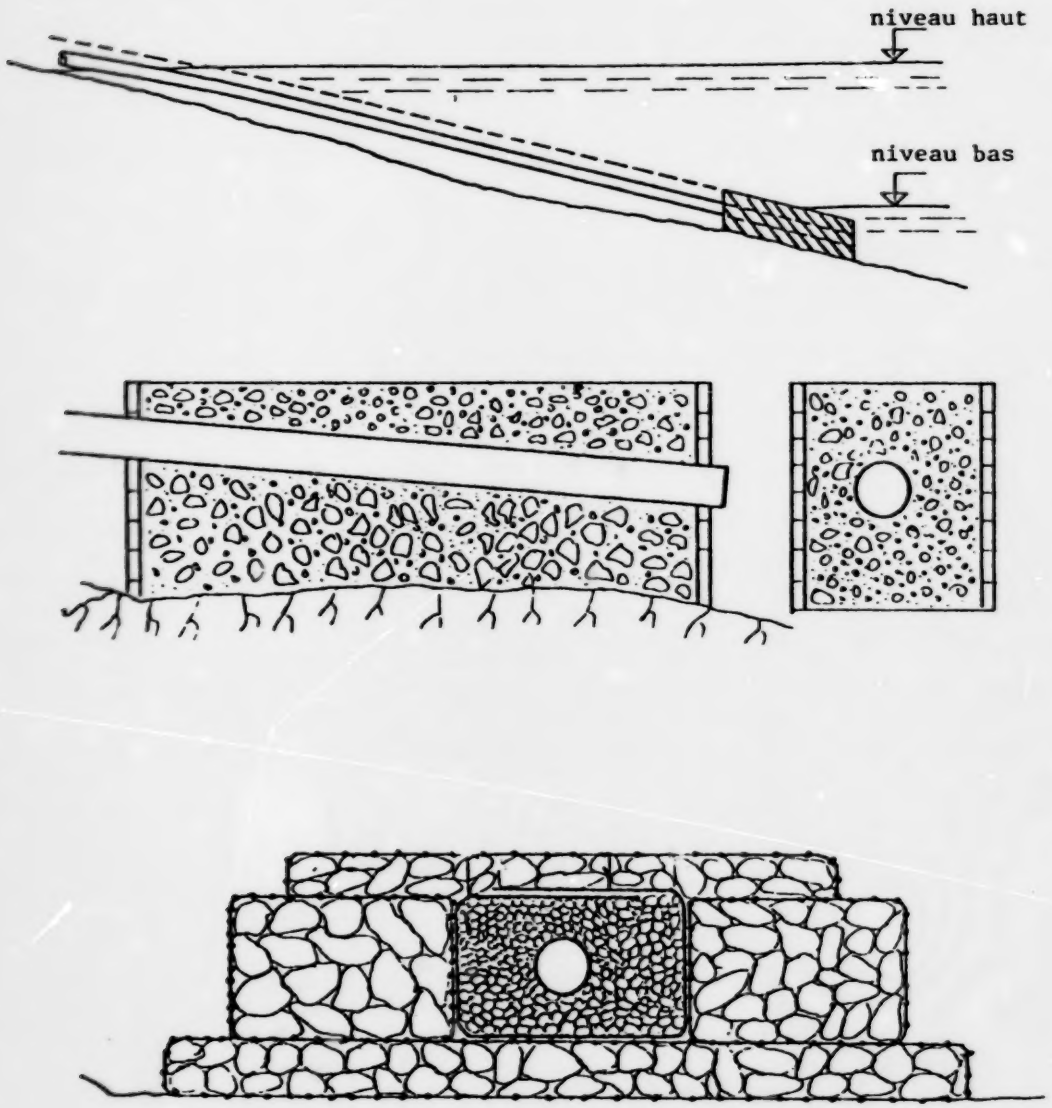
Il convient de connaître, dès la phase de sélection du site, la capacité d'assimilation spatio-temporelle de l'effluent traité du milieu récepteur.

En général, on distingue trois grandes catégories de milieux récepteurs:

1- Sur ou dans le sol. Dans ce cas, le sol assurera une épuration complémentaire et les germes pathogènes seront retenus diminuant ainsi les risques de contamination. À cet égard, les marais et les tourbières sont des milieux très efficaces. Il est important de prévoir un exutoire de type à diffusion afin d'obtenir une dispersion maximale sur le sol.
2- Dans les eaux superficielles. Il faut prévoir un émissaire qui permettra à l'effluent de ne pas longer les rives et d'être dilué le plus possible. Selon notre opinion, tout déversement d'effluent de lagunage dans un lac est à proscrire.
3- Dans le milieu marin. D'après Smith et Tilsworth (1983), l'environnement océanique est très tolérant à un effluent d'eaux usées. Un émissaire en milieu turbulent loin des rives est tout de même nécessaire (figure 5) afin d'assurer une dispersion maximale.

3.8 Désinfection

Lorsque le nombre de bassins est élevé (minimum de trois bassins avec une superficie maximale de 5 ha/bassin) et le temps de rétention assez long (de l'ordre de 6 mois), une concentration de 400 colonies/100 ml de coliformes fécaux peut être atteinte en été. Dans ces conditions, un système de désinfection n'est pas requis.



(tirée de Rosendahl, 1983)

FIGURE 5. ÉMISSAIRE MARIN

3.9 Coagulant

Parker et Uhte (1975) révèlent que l'alun appliqué à l'aide d'un bateau à moteur sur de petits étangs deux fois par année, a permis de diminuer les MES à une concentration inférieure à 10 mg/ℓ. En fait peu d'expériences ont été menées sur l'utilisation de coagulants dans le lagunage naturel en région froide.

CONSTRUCTION ET ENTRETIEN

4.1 Construction

La mise en oeuvre du lagunage repose essentiellement sur la réalisation de bassins en terre. On peut faire appel aux techniques normales de construction utilisées en zone tempérée lorsqu'on n'est pas en présence de pergélisol.

Dans les zones de pergélisol, il est nécessaire de tenir compte, lors de la conception, des conditions spéciales de ce milieu. Dans le cas de sols riches en glace, il faudra faire appel à des entrepreneurs spécialisés avec ce milieu.

4.1.1 Étanchéité des étangs. Il fut un temps où la construction d'un étang était fondée sur le concept d'une étanchéité rigoureuse. Ce critère d'étanchéité est trop sévère et trop coûteux.

Si l'on recherche un étang étanche, le coefficient de perméabilité (K) du sol servant aux étangs devrait être égal ou inférieur à 10^{-5} cm/s. En pratique, le premier bassin qui reçoit les eaux usées brutes, voit l'addition constante des matières solides contribuer à l'étanchéité par colmatage, rendant ainsi ce basssin peu perméable. Les autres bassins n'ont pas à être aussi étanches à moins qu'il y ait un risque réel de contamination des sources présentes ou futures d'alimentation en eau potable.

Il est possible d'imperméabiliser les étangs de façon à réduire les pertes si une telle exigence est requise. Une méthode consiste à ajouter des particules fines (silt, argile, bentonite, etc...) ou encore à traiter les sols avec des matériaux imperméabilisants. Le recours à une membrane imperméable est une méthode simple mais très coûteuse qui devrait être évitée.

4.1.2 Détails. Certains détails qui ont fait l'objet d'une attention particulière peuvent éviter par la suite de nombreux problèmes; en voici quelques-uns:

1- Lors du dégel printanier, la durée du ruissellement a tendance à être courte et le débit très violent dans les bassins hydrographiques septentrionaux; on doit alors prévoir des fossés d'évacuation et de drainage qui interceptent le ruissellement.
2- Les routes et les chemins d'accès ne doivent pas être situés là où il peut y avoir des accumulations de neige ou un englacement.
3- Une échelle graduée permettra de mesurer facilement et rapidement la hauteur du plan d'eau.
4- L'enrochement est un moyen efficace pour une protection anti-batillage.
5- Une pente de fond de 0,5% vers le moine de pisciculture permettra une vidange complète du bassin lorsque le besoin s'en fera sentir.

5- Il convient de prévoir l'installation d'un débitmètre à l'affluent et à l'effluent.
6- Toutes les personnes qui travaillent à l'exploitation devront être vaccinées contre les principales maladies d'origine hydrique.

4.2 Entretien

La bonne marche des installations et leur longévité sont directement liées à l'intérêt de l'opérateur et à son acceptation des responsabilités. L'idéal serait que l'opérateur et son remplaçant soient choisis dès le début de la construction de façon à ce qu'ils reçoivent une formation sur place tout au long des travaux. Un manuel d'exploitation et d'entretien bien rédigé devrait leur permettre de résoudre les problèmes les plus fréquents et de rejoindre facilement des spécialistes lors de problèmes inédits.

4.2.1 Curage des boues. Dès le stade de la conception, une politique de récupération des boues devra être bien établie. La destination des boues ainsi que son mode d'extraction seront choisis en fonction des moyens disponibles sur les lieux. Deux méthodes sont ordinairement utilisées pour l'extraction, soit:

- pelletage des boues après séchage du bassin;
- pompage des boues sous forme liquide.

La première alternative est pratiquement limitée aux étangs constitués d'un minimum de trois bassins pour qu'il y ait toujours deux bassins en service l'été.

Au sujet de la fréquence du curage, Guerrin (1981) estime qu'on ne peut fixer de terme tant cette accumulation des boues est variable puisqu'elle est fonction:

1- de la charge organique traitée;
2- de la nature du réseau d'égout domestique;
3- des conditions climatiques influençant la prolifération des algues, la vitesse de stabilisation des boues, etc.;
4- de la profondeur du bassin considéré.

Le piège à boues devrait permettre l'accumulation des boues pour une période de 4 à 5 ans.

CONCLUSION

De prime abord, il semble que le lagunage soit une technique qui demande des études approfondies et coûteuses lui faisant ainsi perdre son avantage de rusticité et de simplicité. C'est pourquoi l'étude de la conception d'une lagune doit être adaptée aux conditions qui la gouvernent. Les principaux objectifs imposés au concepteur sont la préservation de la santé publique et la protection de l'environnement. Des études de topographie, de géologie, de géotechnique et d'hydrogéologie peuvent alors s'avérer nécessaires pour atteindre ces objectifs.

Même si la "règle d'or" en matière de conception est la rusticité, il n'en demeure pas moins que le rendement dépend essentiellement de la conception de l'étang. Ce dernier devra être constitué de 2 à 3 bassins en série d'une profondeur variant de 1,0 à 1,50 m en été et ayant une charge globale

maximale de 22 kg DBO_5/ha x d. On préconise un minimum de deux conduites à l'affluent et un régime hydraulique favorisant l'écoulement en piston sauf pour le premier bassin dans lequel un mélange complet est désirable.

L'exploitation et l'entretien des bassins, même s'ils ne demandent pas une main d'oeuvre spécialisée, doivent être confiés à des personnes intéressées. Ces personnes seront formées sur place et devront posséder sous forme de documents toutes les explications des opérations à réaliser.

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APPENDICE

EXEMPLE DE CALCUL DE DIMENSIONNEMENT

Nombre d'habitants : 800 hab
Débit moyen (60 jours) : 375 ℓ/hab x d
DBO_5 moyenne à l'affluent (60 jours): 180 mg/ℓ

1) Calcul de la charge

$$800 \text{ hab} \times \frac{375 \text{ ℓ}}{\text{hab} \times \text{d}} \times 180 \frac{\text{mg}}{\text{ℓ}} = 54 \text{ kg } DBO_5/\text{d}$$

2) Surface du premier bassin

La charge maximum étant de 30 kg DBO_5/ha x d, la surface requise est:

$$\frac{54 \text{ kg}}{\text{d}} / \frac{30 \text{ kg}}{\text{ha} \times \text{d}} = 1{,}8 \text{ hectares}$$

3) Surface du deuxième et troisième bassins:

la charge enlevée au premier bassin est: $75\% \times 54 \frac{4 \text{ kg}}{\text{d}} = 40{,}5$ kg/d

La charge restante est: $(54 - 40{,}5) \frac{\text{kg}}{\text{d}} = 13{,}5$ kg DBO_5/d

La charge maximum étant de 5 kg DBO_5/ha x d, la surface requise par bassin est:

$$\frac{13{,}5 \text{ kg}}{\text{d}} \times \frac{\text{ha} \times \text{d}}{5 \text{ kg}} \times \frac{1}{2} = 1{,}35 \text{ hectares}$$

4) Superficie totale:

1,8 ha + 2 (1,35 ha) = 4,5 hectares

Ainsi le premier bassin occupe une superficie de l'ordre de 40% de la superficie totale.

5) Charge moyenne:

$$\frac{54 \text{ kg}}{\text{d}} \times \frac{1}{4{,}5 \text{ ha}} = 12{,}0 \text{ kg } DBO_5/\text{ha} \times \text{d}$$

APPENDICE (suite)

EXEMPLE DE CALCUL DE DIMENSIONNEMENT

6) Emmagasinage d'hiver

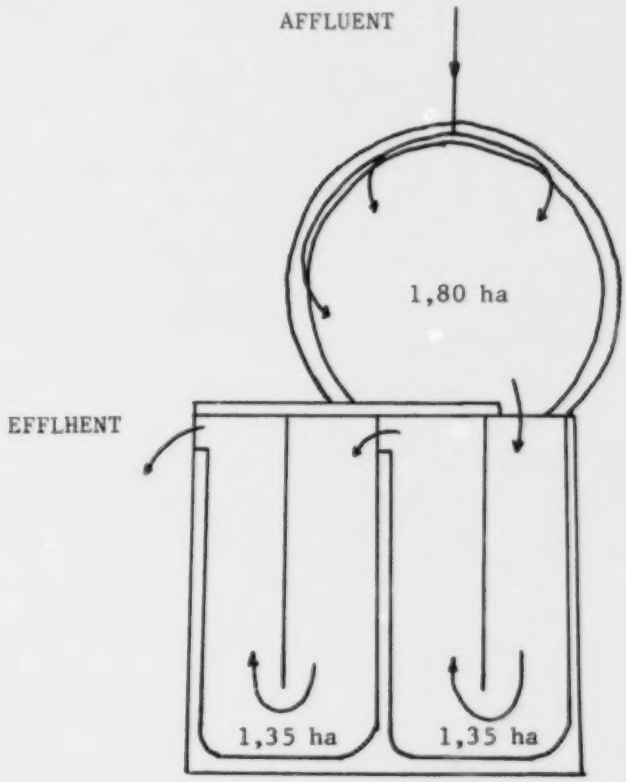
La durée de l'emmagasinage est de 6 mois (1er novembre au 1er mai)

Volume d'eau à emmagasiner: 800 hab x 180 d x $\frac{0{,}375\ m^3}{hab \times d}$ = 54 000 m^3

Le volume d'eau à emmagasiner sera réparti sur les 3 bassins, ainsi la profondeur angmentera de: $\frac{54\ 000\ m^3}{45\ 000\ m^2}$ = 1,20 m

En considérant une hauteur d'eau de 0,75 m dans les bassins avant le gel, la hauteur d'eau maximale sera de 1,95 m au printemps.

7) Un schéma de conception est proposé à la figure 6.



longueur / largeur : 5 / 1

FIGURE 6. SCHÉMA DE CONCEPTION

THE WATERLOO CLARIFIER - PRINCIPLES AND PRACTICE

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ABSTRACT

Research in our laboratory on the fundamentals of clarification over a 10 year period have led to the viewpoint that density fluctuations or internal waves in clarifiers are responsible for removal efficiencies well below those theoretically possible. It has been found that these naturally initiated fluctuations can be controlled by baffles designed to function as wave traps.

Experiments carried out in 4-ft and 8-ft diameter model circular clarifiers show that the baffles reduce low frequency wave amplitudes and substantially improve the solids removal. This paper discusses briefly the principles of density fluctuations, experimental observations and the reduction of our research results to a new, high performance clarifier design - the Waterloo clarifier.

SOME PRINCIPLES OF STRATIFIED FLOW

The presence of suspended solids in moving liquids leads to unique patterns of flow behaviour. These have been known for over a 100 years, but have not been applied to clarification. In addition to the possibility of internal waves, distortion of the flow lines in front of bluff objects and the phenomenon of selective withdrawal may be observed.

Internal waves result from the presence of a vertical density gradient in a clarifier. A vertical density gradient is inevitable if settling occurs. The initiation of waves is explained in Figure 1. Figure 2 shows a trace of the density variations referred to as waves inside a clarifier. The trace was made by placing a small fibre optics light probe/turbidimeter inside a model rectangular clarifier (Cordoba et al., 1979; Silveston et al., 1980). It has been known for many years that internal waves, like surface waves, propagate at relatively low speeds, reflect around a vertical axis and damp out very slowly. Internal waves reflecting off clarifier walls interfere with waves moving toward the wall resulting in large-amplitude waves (the large, relatively infrequent disturbances in Figure 2). Although the details of the mechanism need to be worked out, large amplitude internal waves in clarifiers apparently cause slow, but large-scale mixing which decreases the settling rate. Consequently, reducing the number and the amplitude of the density waves should increase solids removal in clarifiers.

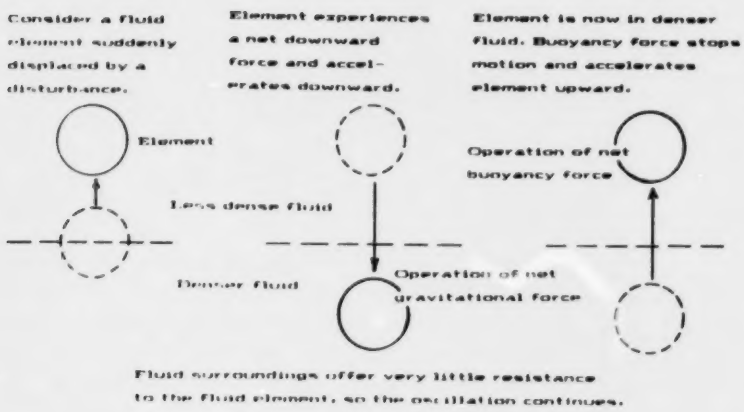

Fig. 1. Why Internal Waves Form.

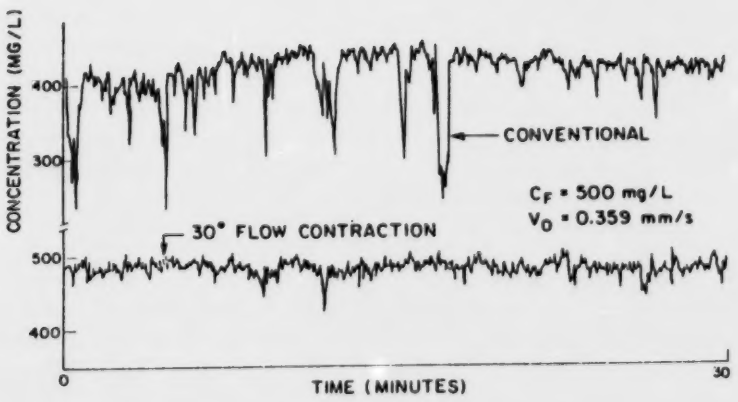

Fig. 2. Traces of Density Variations Inside a Clarifier.

FLOW CONTRACTION CONCEPT

Because the most minute disturbance causes internal waves, preventing wave generation appears to be a hopeless task. Research at Waterloo has concentrated, therefore, on flushing internal waves out of the clarifier, thereby reducing interference and the build-up of large waves. Flushing has been accomplished by creating a plane in front of the clarifier wall where the fluid velocity exceeds the propagating velocity of the wave. Internal waves reaching the wall and reflecting cannot pass back into the clarifier. The local velocity in the constriction, furthermore, is high enough that non-selective withdrawal occurs at the outlet. This means that fluid is drawn from all layers in the clarifier, rather than just the top layer which would be the case for selective withdrawal. The drawback of the latter withdrawal is that it induces large-scale mixing in the clarifier vessel.

The velocity condition just mentioned is achieved just in front of the overflow weir by an upward slanting surface. Figures 3 and 4 illustrate two versions of this surface. In the first (Fig. 3), the surface extends to the clarifier floor. Essentially, the design changes the clarifier wall from vertical to slanted at 15 to 60° to the vertical. The design would be implemented only in new clarifier construction and it is what we refer to as the Waterloo clarifier. The second version is a partial baffle or skirt which does not reach the clarifier floor (Fig. 4). This version represents a retrofit design that could be added to an existing clarifier to improve performance. The figures represent 4-ft diameter, rake-equipped, implementations built in our laboratory.

BACKGROUND

Ever since Camp's ideal clarifier model was proposed some 50 years ago (Camp 1936, 1946), reseachers have sought an explanation for field performance which is often substantially poorer than that predicted from quiescent settling tests. Apart from flocculation effects, short-circuiting or non-ideal residence time distributions were believed to be the source of these deviations. It has been demonstrated, however, that the shape of the residence time distribution, even in well-designed clarifiers, arises from the influent which forms a density current moving along the bottom of the clarifier (Sakata and Silveston, 1974; Silveston et al., 1979). Wouda et al. (1977), find that this situation leads to close agreement with the quiescent settling model so that the deviation source must lie elsewhere.

One reason why settler performance is poorer than that predicted from batch tests may be the presence of internal, density waves in clarifiers. Research published by our group (Cordoba-Molina et al., 1979 A; 1979 B; Silveston et al., 1980) has established the existence of these waves.

THE WATERLOO CLARIFIER

The Waterloo Clarifier is a clarifier whose operation rests on a patented mode of operation (Cordoba et al., 1983), embodying the above

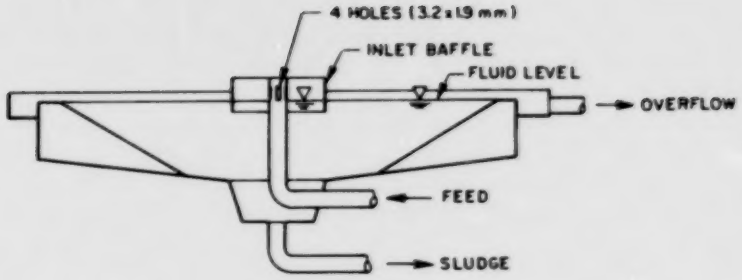

Fig. 3. Schematic (Elevation) of the 4-ft Diameter Clarifier Showing Full 30° Conical Flow Contraction Baffle.

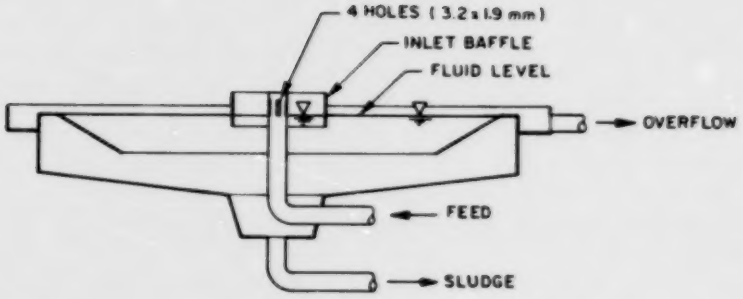

Fig. 4. Schematic (Elevation) of the 4-ft Diameter Clarifier Showing Partial Conical Flow Contraction Baffle.

principles. It incorporates any geometry - circular, rectangular, square, etc. In its design, the vertical walls are replaced by sloping ones. For the circular clarifier, the walls take the form of a truncated, inverted cone (see Figure 5); for the square clarifier, the walls become a truncated inverted pyramid; for the rectangular, only the wall at the outlet is sloped. The angle of the walls of the various geometries lies between 15° and 60° to the horizontal. The slope on the wall must start at a point below the weir such that the Froude number at the region just in front of the weir is 1/ . Effectively, this may be done by arranging the sloping wall to come right to the bottom of the weir, although most of the designs tested in the laboratory use a lip, as shown in the figure. It is likely that even at 60°, surface scraping will be needed, since biologically active wastes tend to form slimes on most surfaces. The surface area is unchanged from that of a conventional clarifier and is governed by recommended overflow rates for different wastes. However, the volume is substantially reduced. This eases foundation requirements for the unit.

The flow contraction concept can also be applied to existing clarifiers by retrofitting a conical baffle or partial baffle as illustrated in Figure 4 (Allen et al., 1983).

PERFORMANCE TESTS

The performance of the clarifier was tested in the laboratory in a 5-ft long scale-model rectangular clarifier and 4-ft diameter circular clarifier illustrated in Figures 3 and 4. These tests were performed using a dilute slurry of diatomaceous earth whose particle size was carefully selected so that the settling rate was the same order of magnitude as that of primary sewage. Overflow rates in the tests were 40 and 80 L/min, corresponding to conventional overflow rates of 300 and 600 US gal/d ft , respectively. Slurry concentration was 500 mg/L. Details of these tests are given by Allen et al. (1983 A). Earlier work which gave similar results has also been published (Murphy et al., 1982).

Clarifier performance was found to be significantly better than that exhibited by the model of the conventional circular design. As may be seen in Fig. 6, at a high overflow rate, flow contraction reduces suspended solids by 18%, while at a low rate, the reduction is 28%. For design purposes, where size is of interest, the reduction in size is remarkable. For example, with suspended solids in the overflow at 28% of feed, adding flow contraction increases capacity by 71%, or, for a constant liquid flow rate, reduces clarifier diameter by 31%.

Experiments shown in Figure 6 were performed with both full and partial baffles. As can be seen, no difference is found.

Experimental results for tests with the rectangular clarifier have been published previously (Cordoba et al., 1979). Results obtained were quite similar to those of Figure 6.

Experiments employing municipal wastewater were performed in Pilot Plant #1 located at the Wastewater Technology Centre, Canadian Centre for Inland Waters, Burlington, Ontario. Raw sewage from the nearby Burlington Skyway Waste Treatment Plant was fed to the pilot plant at 140 L/min into a hydrasieve via a raw feed pump which provided some particle size reduction.

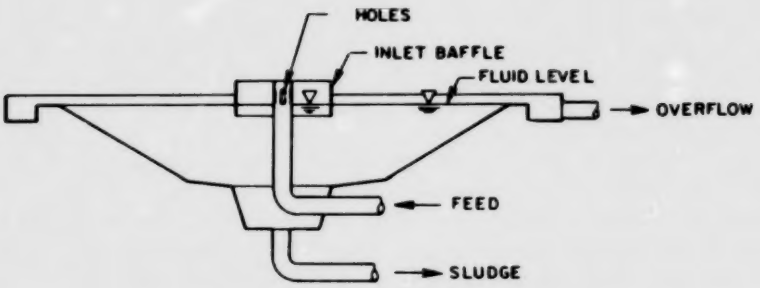

Fig. 5. Schematic of the Waterloo Clarifier.

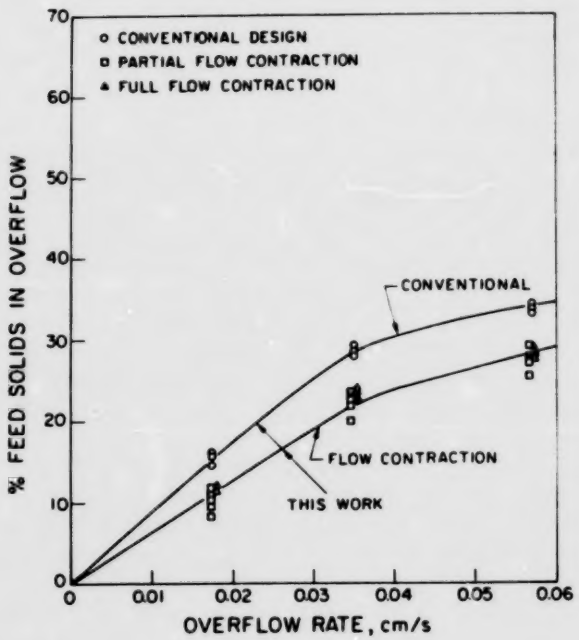

Fig. 6. Experimental Results of the 4-ft Diameter Clarifier.

This plant did not contain a primary clarifier so the influent entered the aerator directly. Clarification was performed in a 8-ft diameter circular clarifier.

Figure 7 is a cut-away drawing of this clarifier. For the experiments, the depth adjustable overflow trough was set at its bottom position so as to give a sidewall depth of 132 cm. Most experiments were carried out with a 60° baffle as illustrated in the figure, constructed of galvanized steel sheet and held in place on a light, angle iron framework. A rake arm was attached to the centre shaft to scrape the baffle. Existing rakes on the floor of the clarifier remained in service with the retrofit baffle design in place.

Two overflow rates were used - 40 and 80 L/min. These translate into the conventional overflow units of 300 and 600 US g/ft respectively. The underflow was held constant at 40 L/min for the experiments discussed in this paper. Flows and turbidities of all streams were continuously monitored, and data were recorded on a mini-computer. Experiments were carefully designed to remove the effects of changing conditions upstream in the Burlington plant. Details are given in the paper by Allen et al., (1983 B).

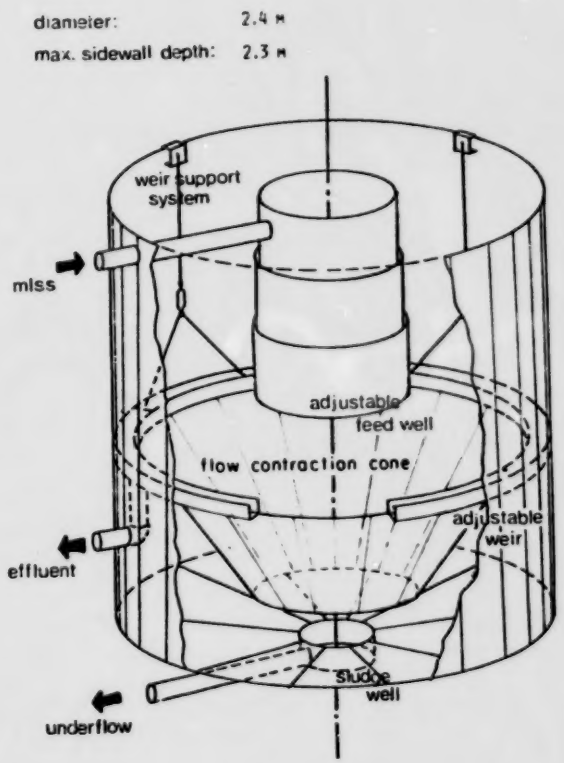

Fig. 7. Cut-Away Drawing of the Circular Clarifier in Pilot Plant No. 1 Showing Position of the Test Flow Contraction Baffle.

The measurements made in the WTC facility with activated sludge are shown in Figure 8. Curves have been drawn through the points corresponding to baffled and unbaffled operation. It can be seen that a remarkable improvement results from using the contraction. The decrease in mean effluent solids, expressed as mg/L, seems to be independent of the overflow rate for the two rates studied. Scales along the bottom give the results as a function of the raw overflow in L/min, as well as in the customary textbook units of US gal/sq ft. A reduction of about 50% in effluent suspended solids is seen at an overflow rate of 40 L/min. The same absolute reduction corresponds to a percentage decrease of 32% at 80 L/min. Results have been corrected to a constant mixed liquor feed concentration of 2.0 g/L.

For the design situation, if an overflow suspended solids of 15 mg/L is to be achieved, the flow capacity can be increased by 27%. Alternatively, the clarifier diameter may be reduced by 13%.

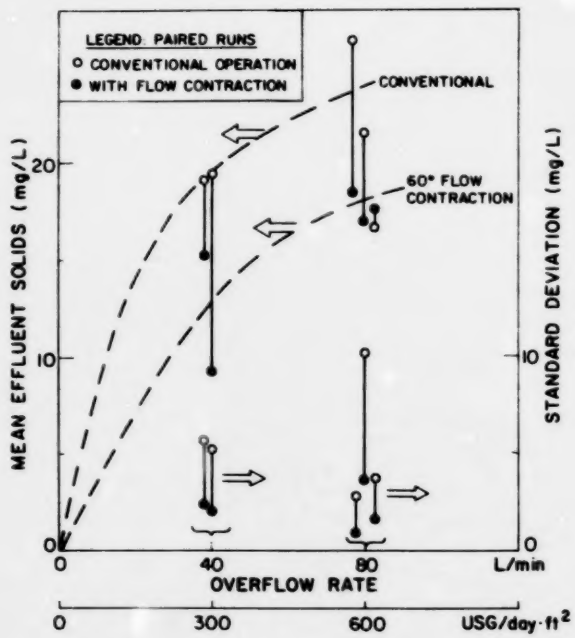

Fig. 8. Effect of 60° Flow Contraction Baffle in Effluent Removal in 8-ft Diameter Secondary Clarifier (Corrected to 2 g/L mixed liquor concentration).

The scum collection provisions in the circular clarifier were very primitive, and some sludge escaped with the overflow. This occurred with the baffle in place and also without a baffle, but was more noticeable in the latter case. To test whether lifted sludge intercepted by the partial baffle contributed to the very favourable results with the baffle in place, the test was repeated using a denitrification inhibitor N-Serve 24E (Dow Chemical Canada). It was found that the performance of the unbaffled clarifier improved considerably, while that of the baffled unit also improved, but to a lesser extent. At 40 L/min, the absolute decrease corresponds to a 50% decrease in effluent solids, whereas at 80 L/min, the decrease dropped to 13%.

Because of the limited data collected, the improvement in capacity just cannot be predicted.

Density waves were seen in all of the tests made. However, variation of the mixed liquor flow rate or the suspended solids made it impossible to compare observations for baffled and unbaffled clarifiers so as to see whether the flow contraction was damping the density waves as was indicated by laboratory experiments performed at Waterloo in the 4-ft circular clarifier with non-flocculating diatomaceous earth as the settling solid.

An important indication of wave reduction, however, is given by examining the variability of the data record, measured as the standard deviation of the overflow suspended solids concentration. These data are contained in Figure 8, which shows that an order of magnitude reduction in variability is found with the flow contraction present. This reduction in variability is to be expected if the flow contraction device is eliminating large-scale density variations.

CONCLUSIONS

Performance tests indicate that the Waterloo Clarifier is capable of suspended solids reductions both in primary clarification (which resembles the diatomaceous earth experiments) and in secondary clarification exceeding conventional clarifiers of the same surface area. Although cost savings should be appreciable with the new design, it needs to be tested in full-scale operation.

ACKNOWLEDGEMENTS

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IMPLEMENTATION OF SEQUENCING BATCH REACTORS FOR MUNICIPAL WASTEWATER TREATMENT

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ABSTRACT

Sequencing batch reactor (SBR) technology is being implemented at various municipal sites in both the United States and abroad. Total life-cycle cost savings, ease of operation and reliability favor this technology at facilities sized up to 19,000 m^3 per day (5 mgd). Batch treatment has inherent advantages over continuous flow processes in many applications.

Current research investigations on SBR's concern controlling sludge settleability, nitrification-denitrification, and biological phosphorus removal.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

INTRODUCTION

This paper discusses progress of full-scale implementation and laboratory research on batch reactors. In the United States and several other countries sequencing batch reactors (SBR) are being designed and constructed for municipal facilities to meet a variety of effluent requirements. Studies are being conducted to seek operational patterns to achieve nitrification, denitrification, and biological phosphorus removal.

Historical technology is replete with examples of batch treatment of municipal wastewater. Sidwick and Murray (1) have outlined the evolution of batch processes into continuous flow processes in England. Early, in land treatment practices, it was recognized that intermittent irrigation of wastewater was necessary for reaeration of the soil to occur. In the year 1897, ways to conserve land were investigated. Vessels were constructed which contained various fine media (3mm) to try to duplicate soil surfaces. The first of these were operated as simple FILL and DRAW in accordance with intermittent land application experience. Later, a cyclic process was instituted which consisted of 2-hour FILL, 1-hour stand full and 5-hour drain. The cycle was repeated twice daily. Seventy-eight percent removal of organic matter was reported. The need to supply oxygen in the proper amount for active biological oxidation was the key finding of this early work. From this point on, two divergent developments led to continuous flow processes. One utilized larger media with greater void space so that wastewater splashing from surface to surface entrained oxygen from the surrounding air. This, of course, is the now familiar biological trickling filter. The other retained the concept of a flooded vessel containing media, but instituted continuous flow with suitable valving and supplied air by an external blower. This technology was the progenitor of the Hays Process (2) and

Imhoff's contact aerators (3). Conceptually, Imhoff discussed a variation of contact aerators which described rotating biological contactors before their actual invention.

The precursor to the various, now familiar, continuous flow activated sludge processes was actually a FILL and DRAW system operated as a batch process. Arden and Lockett (4) in 1914 were among the first to show the benefit of retaining substrate adapted organisms for efficient treatment. Working with 2.3 liter flasks containing Manchester, England, raw wastewater, they showed that the batch aeration period to achieve nitrification could be reduced from 5 weeks to 9 hours if the sludge that accumulated from each batch were retained in the flask after decanting the nitrified liquid. They coined the term "activated sludge" to describe the resultant biological mass. In later full-scale studies these same investigations (5) showed efficient treatment with an 8-hour batch cycle, as shown in Table 1.

TABLE 1. BATCH TREATMENT OF MANCHESTER WASTEWATER (modified from Ref. 5)
Year 1915

Analytical Results, mg/L			Operational Cycle, hours
	Raw	Effluent	
$KMnO_4$ (Oxygen Absorption)	124	18	Filling, 1 Aeration, 4
NH_4-N	37		Settlement, 2 Discharge, 1
Organic-N	12	1.9	
NO_2-N	--	1.4	There were 3 cycles each day and 20 percent activated sludge, by volume, was retained in the reactor after each cycle.
NO_3-N	--	14	

PRESENT DAY BATCH PROCESSES

Many present day industrial processes are batch processes; such as the dairy, steel, pharmaceutical and antibiotic, fine chemical, cosmetic, and paint industries.

In some instances, state regulatory agencies require that toxic wastes, such as cyanide, be detoxified by batch operation to insure the effluent can be monitored before discharge.

Most modern municipal treatment facilities utilize batch treatment in one guise or another, such as anaerobic digestion, aerobic digestion or intermittent discharge lagoons. Tchobanoglous (6) calls attention to the fact that the classical biological oxygen demand test (BOD) is a batch assay; however, the resultant data are used to design continuous flow processes.

Renewed interest in main stream batch treatment resulted from a mentorial article by Pasveer (7). He reported on the conversion of a continuously operated oxidation ditch; treating the waste from Sancta Maria Hospital, Noordwijkerhout, Holland, into a discontinuous discharge process with intermittent aeration. The operational change resulted in the control of a

filamentous activated sludge, improved clarification and yielded 90 percent denitrification.

Gronszy (8) has applied this concept of single tank treatment systems in Australia. Extended aeration processes are operated with continuous inflow but discontinuous discharge. The liquid level in the tank varies and various type decant devices are used. Aeration is provided in a cyclic manner to encourage nitrification and denitrification.

These two approaches provided a high degree of treatment due to the fact that they are both lightly loaded systems. Since the volume of the tank was large in comparison to the influent flow, the possibility of short circuiting influent to effluent was minor.

Irvine, the doyen of current investigations of batch process, has suggested a uniform terminology and united the batch concept with modern control strategies (9).

BATCH COMPARED WITH CONTINUOUS FLOW

Most simply stated, the conceptual difference between batch and continuous flow is that continuous flow processes have spacially related unit operations, whereas unit operations are timed sequentially in batch processes.

The generic batch process can have several modifications, such as single reactor, multiple reactors in series or parallel, and sequencing reactors; all with variable liquid levels. The most simple of these would be a variable volume single reactor as depicted in Figure 1, which would be suitable for a rural or small industrial situation where no flow occurs for part of the day. The figure shows the different time dependent modes that would occur in a complete cycle of a 1-tank reactor. The modes are labeled according to Irvine's (9) recommendation. The volume and cycle time are only illustrative. For any batch system, FILL and DRAW must occur in each cycle, but REACT, SETTLE, and IDLE could be eliminated depending on the objective of the treatment. Alternatively, other modes could be inserted by varying time and operational controls within a cycle. Systems of this nature have a great deal of flexibility. Table 2 compares batch and continuous flow processes. In several instances drastic differences can be noted and their impact on design, operation and effluent quality are apparent.

TABLE 2. COMPARISON OF BATCH AND CONTINUOUS PROCESSES

Parameter	Batch	Continuous
Concept	Time Sequence	Spacial Sequence
Inflow	Periodic	Continuous
Discharge	Periodic	Continuous
Organic Load	Cyclic	Even (by convention)
Hydraulic Load	Cyclic	Even (by convention)

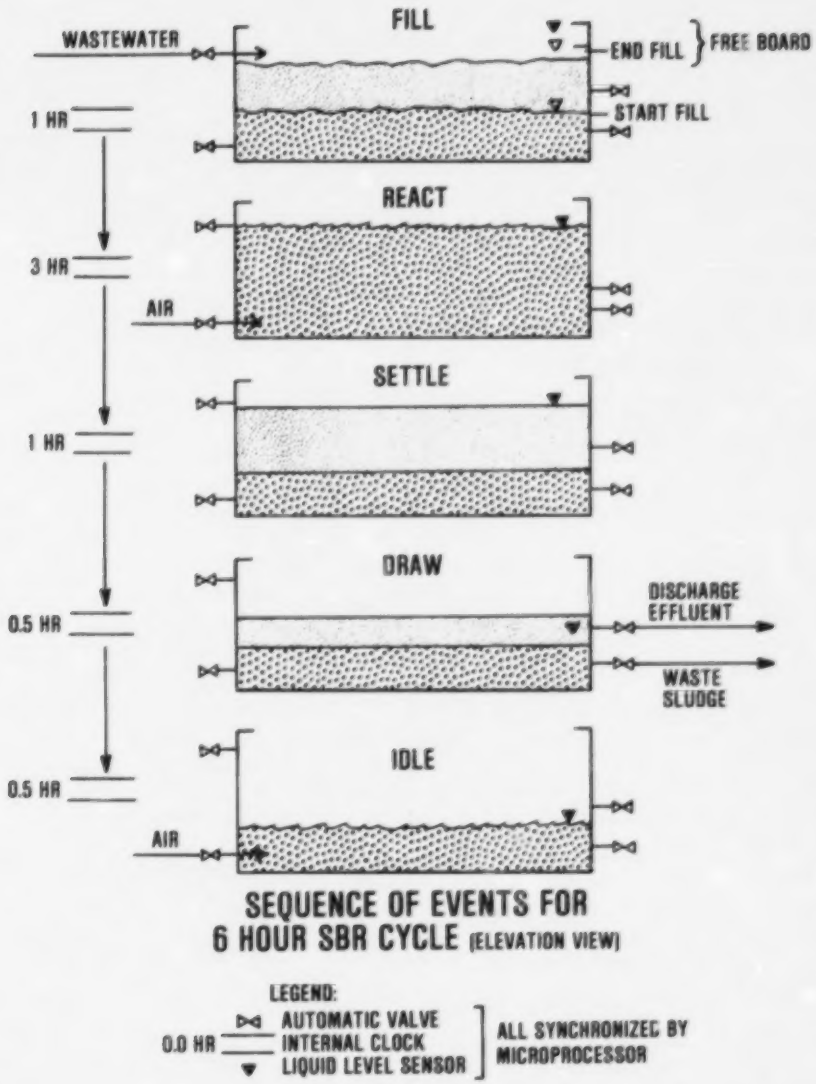

FIGURE 1. SEQUENCE OF EVENTS FOR 6 HR SBR CYCLE (ELEVATION VIEW)

TABLE 2. COMPARISON OF BATCH AND CONTINUOUS PROCESSES (cont'd)

Parameter	Batch	Continuous
Aeration	Intermittent	Continuous
Mixed Liquor	Always in Reactor, No Recycle	Recycles through Reactor and Clarifier
Clarification	Quiescent Hydraulics	Hydraulic Motion
Flow Pattern	Perfect Plug	Complete Mix or Approaching Plug
Equalization	Inherent	None
Flexibility	Considerable	Limited
Hydraulic Sizing	Variable	Uniform

Since the discharge is periodic, the possibility exists, within constraints, to hold effluent until some predetermined specific residual is obtained.

The cyclic organic and hydraulic loading coupled with inherent equalization allow control to be exerted over substrate tension in the reactor by suitably matching oxygen supply.

No recycle pumping is required, since the mixed liquor is always in the reactor, thereby saving on energy. All the active microbial mass is available and not shunted through an inactive period in a clarifier.

Liquid solids separation occurs under near ideal quiescent conditions. During the settling period there is no hydraulic motion due to inflow, outflow or recycle, as in continuous flow systems. The effluent exit piping must be sized larger than influent piping because flow accumulated over a long time span is discharged in a much shorter time span.

In situations where the influent flow is continuous, two or more tanks operated in sequence would be needed to treat the flow. A two-tank sequencing batch reactor (SBR) schematic is shown in Figure 2. As one tank progresses from the FILL to the REACT mode, flow is switched to the second tank, which starts a new cycle in the Fill mode. When capital costs and operational controls are considered, Irvine and Richter (10) indicate a three-tank system is near optimum for flows above 3,785 m^3d (1 mgd).

Because sequencing batch reactor technology is new to both designers and treatment plant operators, the major applications are currently for facilities of 19,000 m^3 per day (5 mgd), or less. This lack of experience on large size facilities is no barrier to implementation. Between the years 1983 and 2000, it is anticipated that 3,000 municipal facilities with design flows less than 7,500 m^3 per day (2 mgd) will need to be constructed in the United States. About 2.5 billion dollars will be expended on these sized treatment plants.

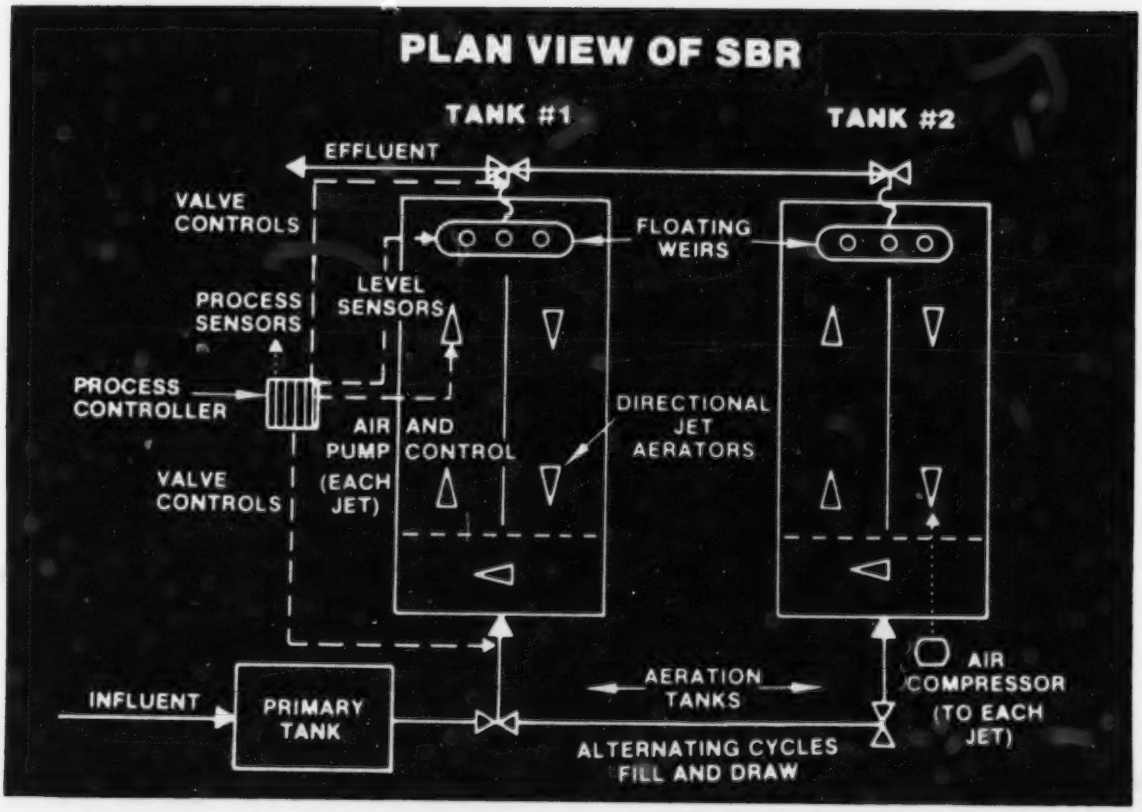

FIGURE 2. SEQUENCING BATCH REACTORS

OPERATIONAL CONSIDERATIONS

Since the majority of the SBR's being constructed are of small size a recent discussion of energy utilization at municipal facilities by Smith is of interest (11). This discussion shows that in the 3,785 m^3 per day (1 mgd) size range energy consumption is only 8.1 percent of the operating cost and labor accounts for 68.2 percent of the cost. Therefore, the SBR process with automated valving, and a process controller which eliminates major labor times, is a logical selection at small facilities. In Australia, where modifications of the SBR have been employed for several years, it is common to have one operator manage four facilities in a several hundred square kilometer area. Also, designers are beginning to recognize the attributes of the SBR in many situations. Barth has noted flexibility, equalization, quiescent settling, elimination of recycle pump, and process reliability, as criteria that direct selection of batch systems for many installations (12).

Bathija has described various SBR tank designs and operational schemes (13). Racetrack, rectangular (length to width ~ 4:1), and circular SBR facilities have been constructed. Recommended side-water depth is 3 to 4.5 m (10 to 15 ft). If both aeration and mixing are independently necessary, a jet type aeration device is preferred. If separate mixing only is not a process requirement any non-clogging or mechanical aeration system can be specified.

Bathija also suggests SBR solutions to two common problem areas in municipal treatment; reduced hydraulic flow during the early design life of a facility, and increased hydraulic flow during storm events. During the early design life period of reduced flow, liquid level sensors can be set at a fraction of tank capacity. In this way, the length of treatment cycles is the same as design, and power is not wasted by over-aeration during an excessively long cycle time. There are several approaches to manage increased hydraulic flow during storm flows. If level sensors are left in their usual position, increased flows will cause the number of cycles per day to increase. If this should cause a reduction in efficiency the level sensors could be raised, within free-board limits, to keep the cycle time near design requirements. Either of these two management approaches could be coupled with modification of the time periods within a cycle. For instance, the time devoted to IDLE could be eliminated and that time period shifted to increase REACT time, yet not alter the overall design cycle time.

The SBR process has an outstanding reliability feature during increased hydraulic flows. The mixed liquor solids always remain in the reactor and cannot be washed out by hydraulic surges. Also, there is no recycle flow from a separate clarifier and, therefore, an SBR process does not rely on return sludge capability to maintain an adequate level of mixed liquor solids during increased hydraulic flow.

EQUIPMENT SELECTION

Because the SBR process is rather new, several types of instrumentation are being pursued. Some designs are employing photoelectric level sensors, others use float level switches for controlling reactor liquid level. Automated valves for control of reactor influent and effluent range from pneumatic devices to motorized gate valves.

Effluent discharge devices can be submerged standpipes, tipping weirs, float supported collector pipes, or submerged pumps.

In all but the smallest systems, commercial process controllers are used to time and control the process sequence. In systems under 20 m^3/day (5000 gpd) simple time clock and electrical solenoid valves would be recommended.

INDUSTRIAL APPLICATION

Dairy Waste

The Springfield Creamery in Springfield, Oregon, produces a waste that has a large concentration of soluble oxygen demand, intermittent flow, and extreme pH fluctuation. Flow is 40 m^3/day (10,000 gpd). An adjoining meat packing plant has a large concentration of suspended oxygen demand, intermittent flow, and stable buffered pH of about 7. Flow is 8 m^3 per day (2000 gpd).

Upon learning of the inherent equalization of an SBR process, ease of operation, wide flexibility of treatment options, and simple design considerations, the two companies plan to mix these waste streams for combined treatment. The SBR will be an earthen basin with dimensions of 7.5 m x 4.5 m x 2.4 m (25 ft x 15 ft x 8 ft). Aeration will be provided by a submerged pump with an air blower exhaust line tapped into the intake side of the pump. Since the flows are intermittent, a single tank system will be adequate.

Chemical Waste

Due to the facts that SBR's can be operated to produce a defined residual concentration of a pollutant with no possibility of short circuiting, and that quiescent settling produces very clear effluent, the process has been selected to treat a combination of landfill leachate and chemical manufacturing wastes in Niagara, New York.

The plant wastes are generated from the industrial operations and the leachate is collected from a remote landfill area via collection underdrains and hauled by truck to the plant site. Both type wastes contain several hazardous organic pollutants. Daily design flow is 600 m^3.

The exsting treatment at the industrial site consits of a storage tank, neutralization tank and carbon absorption columns; with discharge of the column effluent to the city sewer. Operation has shown that with this physical/chemical process, excessive soluble organic matter quickly saturates the absorption capacity of the carbon columns requiring frequent thermal regeneration of the carbon. Upgrading construction will add two biological SBR's, to reduce the soluble organics, and provision for adding ammonia and phosphate to serve as nutrients for the SBR biomass.

Many of the organic components of the mixed waste have slow rates of biodegradation. To insure efficient degradation, the SBR contents will be monitored for total organic carbon during REACT until a predetermined concentration is achieved. Thus, the SBR will decrease the organic loading to the carbon columns and efficient settling will not overly burden the columns with suspended solids. Waste activated sludge from the SBR's will be incinerated to destroy any sorbed hazardous pollutants on the sludge surface.

Pilot SBR operation at a scale of 50 m^3/day (13,000 gpd) has shown a 90 percent reduction of soluble organic carbon from the influent mixed waste. It is therefore anticipated that the useful life of the carbon, before regeneration is necessary, will be increased by almost a factor of 10. Figure 3 is a schematic of the full-scale process as it will exist in 1985.

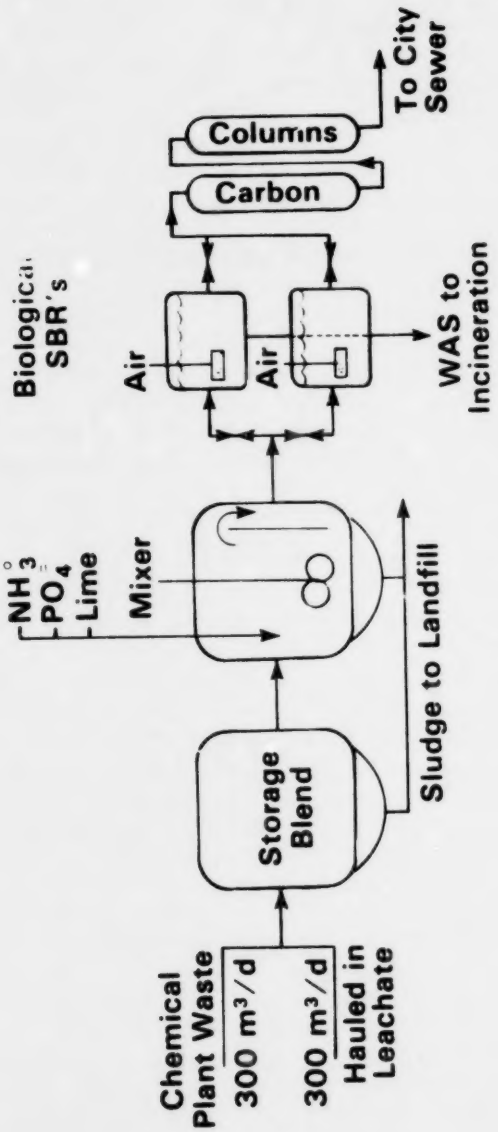

FIGURE 3. SBR-CARBON ADSORPTION FOR INDUSTRIAL WASTE

MUNICIPAL APPLICATION

Recognition of the virtues of SBR's and the emergence of reliable process control hardware lead to the USEPA to fund a development project, conducted by The University of Notre Dame, to re-evaluate municipal batch treatment. The project involved the conversion of the existing 1,500 m^3/day (0.4 mgd) Culver, Indiana, continuous flow activated sludge facility into a two-tank SBR (14). The new flow scheme is the same as shown in Figure 2.

Primary effluent is directed to each tank on an alternate basis as dictated by liquid level sensors in each tank. Pneumatic compression pinch valves control tank switching. Aeration and mixing are provided by directional jet aerators. Effluent is decanted during DRAW by submersible pumps attached to flotation devices which are swivel mounted to allow changes in elevation of the pump suction. All of these functions are programmed into a process controller which operates and monitors all aspects of treatment.

The controller programs for both Tank-1 and Tank-2 operate independently in order to have greater mode flexibility within a cycle. However, a main program monitors the Tank-1 and Tank-2 programs on a time-sharing basis to insure correct sequencing.

The initial evaluation at the Culver facility used an anoxic period for part of the FILL mode and this was programmed into the controller as FILL-1 and FILL-2. During FILL-1, only the mixing action of the jet pumps was activated. During FILL-2, the jet pumps mixed and aearated simultaneously. A typical cycle for the two tanks was as follows:

	Hours in Mode					
	Fill-1	Fill-2	React	Settle	Draw	Idle
Tank-1	1.1	1.8	0.8	0.8	0.7	0.8
Tank-2	1.2	1.9	0.6	0.8	0.6	0.9

The difference in times between the modes of the two tanks is due to flow variation, since the system is controlled by liquid level sensors. At the design flow of 1,500 m^3/d (0.4 mgd) each cycle is completed in 6 hours and both tanks complete 4 cycles each day.

The process conditions at Culver were:

F/M	0.2 kg BOD_5/kg MLSS/d
SRT	20 days
MLSS	2,200 mg/L (70 percent volatile)
SVI	110 ml/g

One year of data and operational experience has been documented for BOD_5 and SS control. Table 3 provides monthly average wastewater and SBR effluent constituents for 8 months of 1980. The City of Culver has an effluent phosphorus limitation of 1 mg/L. To achieve removal of phosphorus, ferric chloride is dosed directly into the primary effluent channel leading to Tank-1 and Tank-2.

Very high efficiency for BOD_5, SS and TP is achieved at Culver with the SBR process. Nitrate nitrogen of about 2 to 5 mg/l is usually present in the

Culver raw wastewater due to run-off from fertilized farm land infiltrating sections of the collection system. Table 3 shows that this influent nitrate is denitrified during the anoxic FILL-1 mode of the SBR.

TABLE 3. CONSTITUENTS OF CULVER, INDIANA PROCESS STREAMS
MAY - DECEMBER, 1980

Location	Milligrams per Liter				
	BOD_5	SS	TP	NH_4-N	Oxidized Nitrogen
Raw Wastewater	173	136	6.3	20	2.8
Primary Effluent	132	81	5.2	19	2.6
SBR Tank-1	8	7	0.4	19	0.4
SBR Tank-2	9	9	0.4	18	0.4
Overall Removal, Percent	95	94	94	5	86

Process conditions to achieve nitrification and denitrification were implemented during a second study period. The anoxic period for FILL-1 was reduced to provide a longer period of dissolved oxygen supply for nitrification. The efficiency of the SBR process during the last five months of 1981 is given in Table 4.

Nitrification and denitrification occurred almost simultaneously, as evidenced by the fact that neither ammonium or nitrate nitrogen reach high concentrations in the reactors. Apparently as ammonium nitrogen is transformed to nitrate, during the low load period at the start of FILL, there is subsequent denitrification due to the high mixed liquor concentration and differential D.O. content of the reactor. When mixed liquor is pumped through the aerated jet nozzle a high D.O., suitable for nitrification is present; then as the mixed liquor circulates as a bulk liquid localized low D.O. conditions can occur.

Once nitrification is established in the Culver SBR, ammonium nitrogen concentration in the reactor can never reach influent levels due to the dilution of influent by the residual reactor contents left after IDLE, and the biological transformation to nitrate. Accordingly, the nitrate concentration cannot reach high levels, and even a low kinetic rate for denitrification would yield fairly efficient nitrogen removal.

TABLE 4. EFFICIENCY OF CULVER SBR FOR NITRIFICATION AND DENITRIFICATION AUGUST - DECEMBER, 1981

Location	Milligrams per Liter				
	BOD_5	SS	TP	NH_4-N	Oxidized Nitrogen
Raw Wastewater	158	150	7	20	2
Primary Effluent	110	82	5	18	2
SBR Tank-1	10	5	0.7	1	1
SBR Tank-2	9	6	0.8	1	1
Overall Removal, Percent	94	96	89	95	-

LABORATORY RESEARCH ON SBR's

Control of Sludge Bulking

Working with 4 L SBR's, fed a soluble synthetic wastewater, Chiesa has correlated sludge volume index (SVI) with aeration time during the FILL period (15). He used the following strategies to manage the substrate tension in the SBR's:

Control Strategy	Unaerated/Unmixed FILL, hr	Aerated FILL, hr	Aerated REACT, hr	SETTLE, DRAW and IDLE, hr
A	2.00	0.00	4	2
B	1.75	0.25	4	2
C	1.50	0.50	4	2
D	1.25	0.75	4	2
E	1.00	1.00	4	2
F	0.00	2.00	4	2

Very spectacular results were obtained as shown in Figure 4. To confirm this observation, an SBR operated under control strategy F, with an SVI of 600 mL/g, was switched to strategy A; and in 15 days the SVI decreased to 100 mL/g.

These researchers explain these observations on the basis of competitive growth of one group of organisms in opposition to other groups. They postulate that strategy A, where substrate tension surrounding the organisms reaches a high level during FILL, and subsequent oxidation of organics and endogenous metabolism during REACT, which places the organisms in periodic starvation, selectively encourages the growth of floc forming organisms.

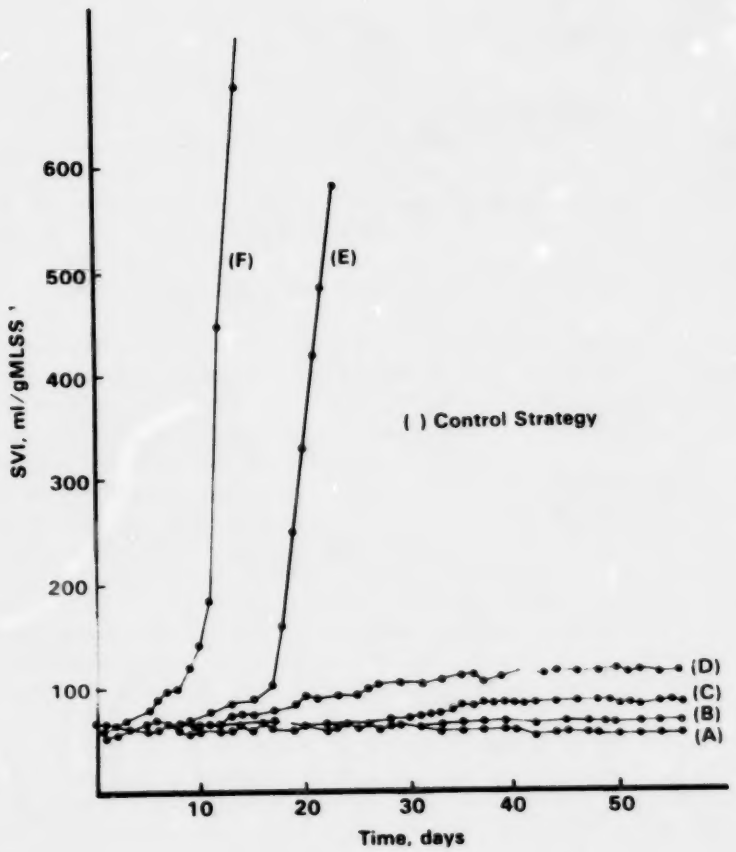

FIGURE 4. SVI AS A FUNCTION OF CONTROL STRATEGY
(FROM REF. 7)

Filamentous organisms survive best when the substrate tension is constantly maintained at a low level such as in strategy F.

The sludge developed under strategy A had a greater specific oxygen uptake rate than did the F strategy sludge; and the A sludge was much more efficient in substrate removal rate than the F sludge.

The authors point out the great flexibility of the SBR process to control substrate tension at any desired level. It can be noted in the above strategies that A would be equivalent to contact stabilization and F would be similar to a completely mixed activated sludge process.

Biological Phosphorus Removal in an SBR

Discussions at the June 1982 "Workshop on Biological Phosphorus Removal" showed a similar process feature for three proprietary phosphorus removal systems (16). This feature is shown conceptually in Figure 5. When activated sludge mixed liquor is subjected to an anaerobic or anoxic environment* with influent organic substrate, there is coincident leaching of phosphorus from the cellular material and sorbtion of organic matter. When environmental conditions are changed to an aerobic state, phosphorus is removed from solution and organics are oxidized.

The following several theories have been proposed to explain this finding:

a) Anaerobic-aerobic staging can result in selection of a biological population that is capable of storing polyphosphates. The actual mechanism for storage and release and the basis of population selection is not known.

b) The regulation of the adenosine diphosphate ratio to adenosine triphosphate (ADP/ATP) controlled by the reversible enzyme polyphosphate kinase has been suggested as a mechanistic factor. When cells are placed in a "stress condition" such as lack of oxygen during an anoxic period, large amounts of polyphosphates are then stored in the cells when subsequent aerobic conditions are provided and cellular growth is rapid.

c) Another theory suggests that polyphosphate storage ability of certain bacteria allow them to compete and survive in an alternating anoxic-aerobic environment. The energy derived by hydrolysis of polyphosphate (ATP) may be used directly for transport of substrate across the cell membrane in the anoxic condition.

d) Polyhydroxy butric acid (PHB) has also been suggested as an important consideration. It has been noted that PHB is formed during anoxic conditions, and organisms such as Acinetobacter can accumulate both PHB and phosphates.

These various mechanistic theories seem to show that some type conditioning occurs during the anoxic period; whether enzymatic hydrolysis, uptake of a metabolite or environmental stress that confers a competitive advantage to certain phosphate storing organisms when conditions are switched to an aerobic state.

*The term anaerobic refers to environments which have no measurable concentrations of either oxygen or oxidized nitrogen. Often times appreciable concentrations of carbonaceous materials must be present to maintain this condition. Anoxic refers to environments which have no dissolved oxygen but have oxidized nitrogen present. However, these two terms were used almost interchangeably in the workshop discussions (16).

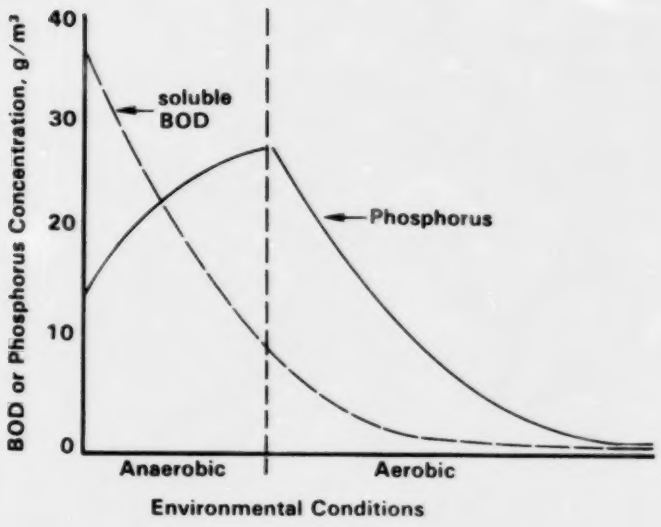

FIGURE 5. BIOLOGICAL PHOSPHORUS AND BOD_5 REMOVAL DUE TO ANAEROBIC-AEROBIC CONTACTING (REF. 17)

The PhoStrip process* utilizes this anaerobic or anoxic leaching to remove phosphorus from a portion of mixed liquor in a side stream stripper tank. Both the A/O** and Bardenpho*** processes remove phosphorus by baffling the biological reactor to maintain the mainstream flow environmental conditions suitable for each process' operation.

Also, all three processes were reported to be subject to inhibition of phosphorus leaching during the anaerobic or anoxic phase if nitrate ion was present. If nitrate was present, provision for denitrification before phosphorus leaching had to be provided.

All of these applications for phosphorus removal are for continuous flow systems. The Municipal Environmental Research Laboratory is employing an SBR for studying biological phosphorus removal because the batch approach should allow close control of environmental conditions. A schematic of the 4 L bench-scale reactor used for these studies is shown in Figure 6. A vast number of operational schemes can be employed using a microprocessor controlled SBR. In order to achieve the release and uptake of phosphorus noted in Figure 5, the following cycle scheme was employed:

Cycle Event	Time	Purpose	Note
FILL+	1 hr	Add substrate	No aeration
STRESS	4.5 hr	Cause phosphorus leaching and denitrification	Mixer on No aeration (Time overlaps with FILL time)
AERATE	1 hr	Cause phosphorus uptake, and oxidize organics	Aeration
WASTE SLUDGE	4 sec	Remove phosphorus from reactor, and control SRT	Aeration
SETTLE	15 min	Form clear supernatant	No aeration
DRAW	5 min	Discharge effluent	No aeration
IDLE	10 min	Adjustment of cycle length	No aeration

*Biospherics, Inc., 4928 Wyaconda Rd., Rockville, MD 20852
**Air Products, Inc., Box 538, Allentown, PA 18105
***EIMCO-PMD, 3839 S. West Temple, Salt Lake City, UT 84115
+Primary effluent from Cincinnati's Muddy Creek treatment plant used.

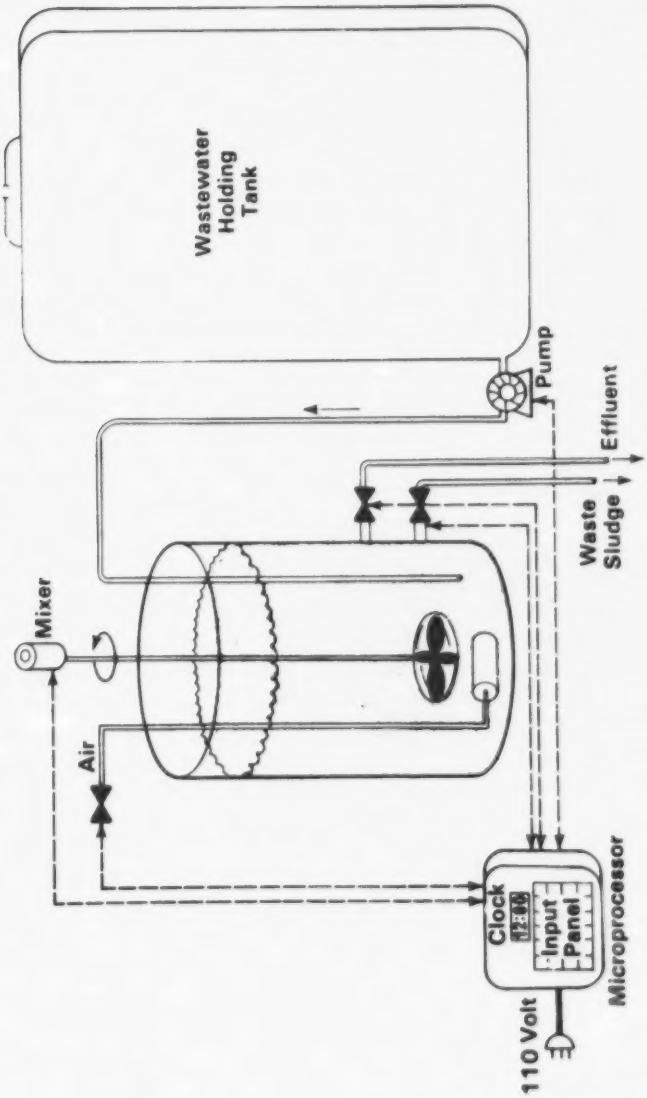

FIGURE 6. 4 L BENCH-SCALE SBR

The total cycle time was 6 hours and therefore the SBR emptied and filled 4 times each day.

The results achieved are shown in Figure 7. During FILL the phosphorus concentration in the reactor is affected by the phosphorus content at the end of the last cycle event (either DRAW or IDLE), the concentration in the feed entering the reactor and the leaching of phosphorus from mixed liquor solids. In the case depicted, the residual phosphorus in the SBR was diluted as primary effluent, with a smaller concentration entered, and a gradual increase to 33 g/m^3 occurred as mixed liquor cellular leaching occurred. During the STRESS period a plateau concentration existed, until aeration was commenced, and cellular uptake reduced the concentration to 2.5 g/m^3. Effluent filtered orthophosphorus concentrations approaching 0.0 g/m^3 have occasionally been noted.

Nitrate nitrogen remained at about 1 g/m^3 during STRESS due to the occurrence of denitrification. A dramatic increase of nitrate formation occurs once aeration is commenced and dissolved oxygen increases.

The SBR was operated under this regime for 30 days (120 cycles) using the Muddy Creek primary effluent. SBR effluent residual orthophosphorus was routinely between 2.5 and 0.5 g/m^3. Influent total phosphorus varied between 20 g/m^3 and 15 g/m^3, with orthophosphorus between 20 g/m^3 and 6 g/m^3. Muddy Creek wastewater is primarily domestic waste.

The process management to control biological phosphorus removal can be illustrated by referring to Figure 7. During FILL and STRESS, conditions have to be provided to cause denitrification of any nitrate and insure that leaching from the cellular mass occurs. STRESS periods ranging from 2 to 6 hours have been studied. Three hours appears to be the minimum time that can be employed. An aeration period of at least 1 hour is necessary to reincorporate the leached phosphorus and oxidize organics. Aeration times greater than this lead to less efficiency due to lysis of cells during endogenous respiration. Wasting of sludge from the SBR must be done when the biological solids have the highest content of phosphorus. This would be at the end of AERATE, which yields a dilute waste activated sludge of the same solids content as the mixed liquor. Or, wasting can be done near the end of SETTLE, which provides a thickened waste activated sludge. Caution is required here in case low dissolved oxygen in the sludge blanket causes leaching of phosphorus. Thus, overall efficiency of biological phosphorus removal dictates that SETTLE time be as short as possible, yet provide good solids separation. A minimum of 15 minutes and a maximum of 30 minutes has been satisfactory.

Time devoted to DRAW is not critical to biological phosphorus uptake directly. Mainly, on full-scale SBR's this time period would be related to effluent weir design to insure proper hydraulics within the reactor, and hydraulics of discharge appurtenances. Provision for an IDLE period is optional in SBR operation for biological phosphorus removal. A short IDLE period, either quiescent or aerated would have no impact on overall efficiency. IDLE is simply a convenience of adjusting cycles between reactors if two SBR's are operated alternately. The same SBR process cycle described above was employed with wastewater from Cincinnati's Little Miami facility. This waste contains a major fraction of wastes from a variety of industrial processes. Consistent biological phosphorus removal could not be achieved. Over a three month period variations to FILL, STRESS, and AERATE were attempted

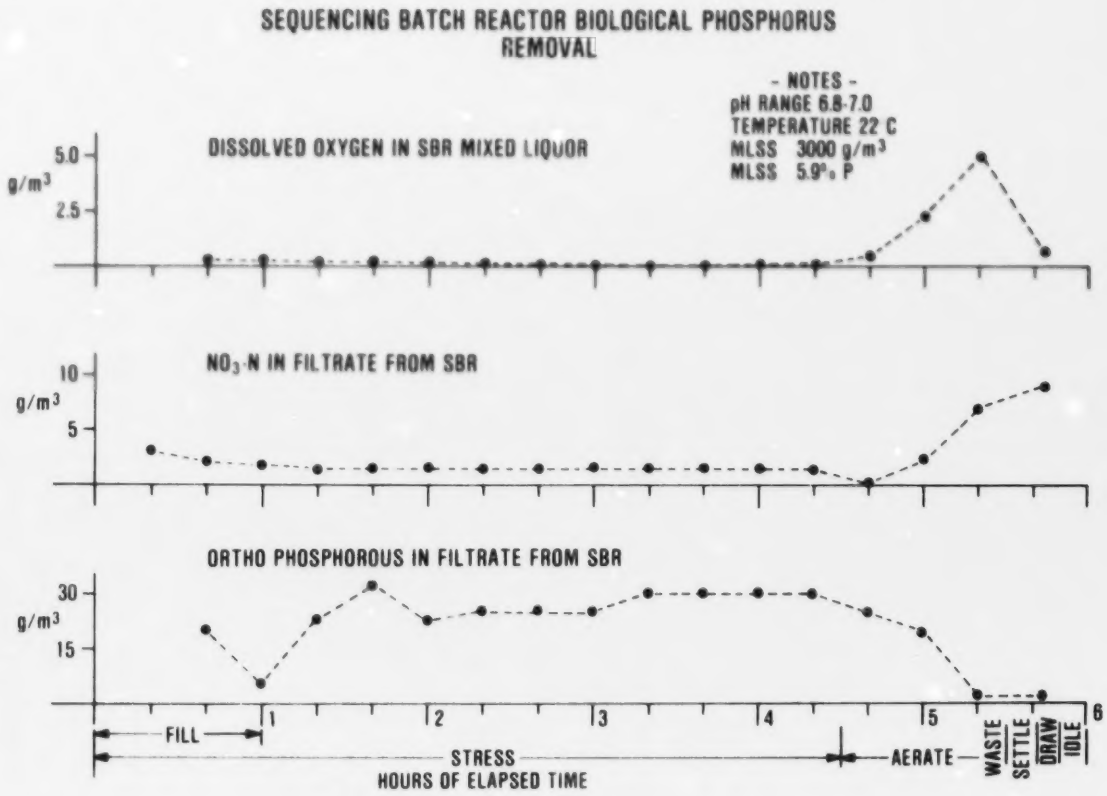

Figure 7. SBR BIOLOGICAL PHOSPHORUS REMOVAL

without success. Efficient removal with residual orthophosphorus values of 0.5 to 1.0 g/m^3, would be noted for several days; then deterioration of efficiency would occur.

These SBR studies, on two wastewaters, have not as yet revealed the controlling parameters to insure consistent, highly reliable biological phosphorus removal. This same conclusion was noted for the three continuous flow biological phosphorus removal processes dicussed at the Annapolis workshop (16).

Presently it would be recommended that any facility plan for biological phopsphorus removal be guided by pilot plant studies prior to final selection. Additionally, in those cases where a high reliability of phosphorus removal is necessary, it would be recommended that a chemical addition system be installed as backup equipment.

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APPLICATION OF BIOLOGICAL BATCH REACTOR TECHNOLOGY IN WASTEWATER TREATMENT

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ABSTRACT

Even though biological batch reactors preceeded the conventional activated sludge process, the use of biological batch reactors had until recently been relegated to the laboratory for process design data generation. Their main function was to provide kinetic data. Only in recent years have batch reators been used at full scale in a variety of operational modes. While the concept of using batch reactors for seasonal industries would appear to be most attractive, to date most applications have been restricted to the treatment of domestic wastewaters.

This paper introduces the concept of applying batch reactor technology, describes its role, highlights experiences in some countries and reports on its first Canadian application in a process retrofit situation.

Data from an intensive monitoring program in Canada illustrates process insensitivity to highly variable organic and hydraulic loads by producing a 5/5 effluent from a single reactor system. Without any attempt at process optimization effluent nitrogen concentrations were 14 mg TKN/L, 2.5 mg NH_3-N/L and 31 mg NO_3 - N/L. Reductions in total phosphorous without use of chemical precipitants exceeded 50 percent.

Experiences confirm that the proper application of biological batch reactor systems is simple, inexpensive, requires a minimum of operator skill and attention yet produces an effluent with characteristics normally associated with more complex biological waste treatment processes.

INTRODUCTION

The selection of an appropriate technology to treat a particular waste, be it of municipal or industrial origin, is based on finding the least cost solution to meet a set effluent target. The conventional effluent parameters are in most instances oxygen demanding substances, as expressed by the biochemical oxygen demand (BOD_5) and suspended solids (SS). Recently, the concern about eutrophication has resulted in setting effluent discharge targets for total phosphorus. That is where phosphorus has been identified as the limiting nutrient. Water bodies may also have a limiting capacity for effluent nitrification thus requiring that nitrification be practiced at the waste treatment plant. On the other hand, if raw water for potable water supplies is abstracted from sources high in nitrate, denitrification may have to occur before effluents are discharged to these sources.

Of late, additional concerns over and above the presence of bacteria and viruses in effluents have been directed at metals and synthetic organic compounds.

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In the early seventies one school of thought proposed that in view of ever increasingly stringent effluent requirements only physical-chemical treatment technology will be able to meet these demands. Biological treatment was going to be relegated to an 'also ran' technology. Thus physical-chemical treatment was hailed by many as the 'in' technology. The advent of computer assisted operation of treatment plants further reinforced this school of thought, since the prevailing school of thought was that it is much easier to automate physical-chemical processes than biological processes. Numerous studies have shown that to achieve the same effluent quality as a biological process, physical-chemical treatment technology is 2 to 3 times more energy intensive than biological treatment. With the advent of the energy crisis this was bad news for the proponents of physical-chemical treatment.

A further consequence of high energy prices was a concerted effort in research and development to modify and enhance biological processes. The objective was to be able to achieve effluent targets normally associated with physical-chemical processes. This has been achieved to a very large measure. Today we have an array of biological processes which give us effluents with low BOD_5, SS, nutrients (phosphorus, nitrogen), metals, and synthetic organic compounds. To a great extent this is attributable to a much improved understanding of the biological, microbiological, physical and biophysical interactions in biological waste treatment.

In this seach for more cost effective technology old processes such as batch reactor technology (1) were re-examined and some new ones such as the two-stage Absorption-Biodegradation process (AB), (2) emerged. It may come as a surprise to some to hear that the so common activated sludge process had its origin in fill-draw sytems dating back to 1893. Pasveer (3) revived the batch reactor concept in the '50s with his original version of the oxidation ditch operated in the fill-draw mode.

BATCH REACTORS

Thus, batch reactor technology represents nothing new, rather something old. Batch reactors though, have been used in the laboratory to investigate waste biodegradability and establish the process kinetics prior to running continuous-flow experiments for process design data generation. We have now come to the stage where batch reactor technology is applied on a larger scale.

Batch reactor systems are attractive because they counter normal concerns associated with small municipal wastewater treatment systems such as:

1. erratic hydraulic and biologic loads;
2. inefficient sludge return (air-lift systems offer no sludge return control);
3. problems with clarifiers due to erratic hydraulic loads;
4. limited available operator skill.

As well, batch systems are less susceptible to sludge bulking than continuous flow systems (4). For batch reactors process operations are reduced to two considerations:

1. mixed liquor settleability;
2. completion of reactions.

Figure 1 compares the conventional activated sludge flow diagram to the batch reactor, and illustrates its compactness and the integration of the biological reactor with the settler.

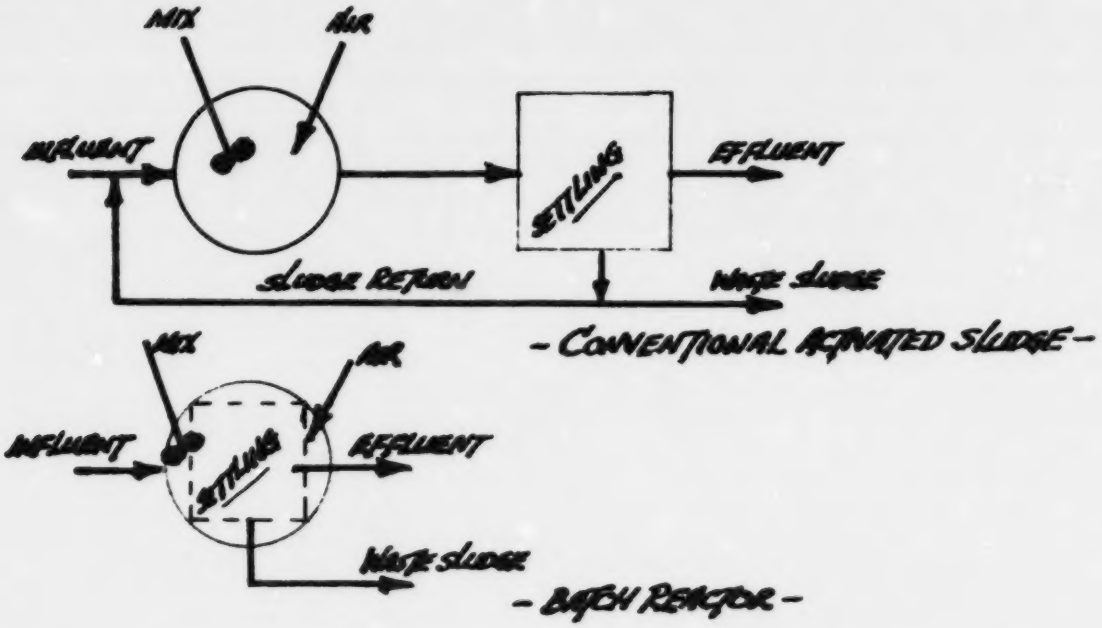

Fig. 1. Conventional activated sluge to batch reactor comparison.

A distinct advantage of batch reactor systems is that it permits you to keep the reactor contents until the previously set effluent criteria is achieved. There is a small problem associated with this though. Facilities to store incoming waste must be available.

Batch reactors are operated on a fill-aerate-settle-decant cycle where liquid/solid separation and biological reactions are integrated within one vessel. A discontinuous system such as a batch reactor, permits settling to occur under almost ideal conditions. In the resulting quiescent environment no new energy is imparted to the liquid. Internal eddies and density currents drop to zero velocity.

PROCESS ADVANTAGES

Batch operated waste treatment systems offer a number of advantages such as:

1. reactor also serves as an equalization basis;
2. organic load is handled by MLSS control;
3. hydraulic load is handled by liquid level control;
4. elimination of sludge recycle;
5. no concern regarding peak hydraulic loads and settling;
6. settling occurs under nearly perfectly quiescent conditions;
7. operationally simple;
8. infrequent sludge wasting (based on settling test);
9. low capital cost;
10. good application for retrofitting existing plants;
11. ability to control effluent quality (a plug flow reactor);
12. operation can be fully automated with micro-processors.

PROCESS DESIGN

The design of batch reactor systems is dependent on the origin and characteristics of the waste to be treated. From a process standpoint batch bench scale reactors are operated in the laboratory to provide information on waste biodegradability and process operation. The various time increments required for each of the fill-react-settle-decant cycles are established. A significant advantage in batch reactor process design consists of the ability to transfer laboratory results directly to full-scale without having to consider the question of scale-up. One point must be kept in mind, especially in a process retro-fit situation, because the effluent will be discharged over a short period of time, effluent piping must be larger than influent piping.

PROCESS APPLICATIONS

Applications would appear to be unlimited. Situations where highly variable and periodic organic and/or hydraulic loads are encountered are prime candidates for application. This does not only apply to small communities, resorts, schools, institutions and camps but should be of specific interest to those industries which generate biodegradable wastes. Easy construction, start-up and simplicity in operation offers batch reactor technology as an ideal solution to seasonal operations. The canning and fruit and vegetable industries are two industrial examples, resorts which operate on a seasonal basis are further examples for batch reactor application. Construction camps fall into the same category.

Low capital cost, simple materials of construction and ease of operation make the batch reactor approach an attractive proposition for developing nations (5).

The number of batch reactor applications is increasing steadily. In the USA installations range from 1 900 m^3d^{-1} to 19 000 m^3d^{-1} in size (6). Batch reactors in conjunction with conventional physical-chemical processes have also been used in an industrial application where a zero discharge requirement had to be met. Effluent reuse and recycle was practiced (7).

In Australia, numerous applications of batch reactor technology for municipal waste treatment have been found to be cost-effective solutions. Investigations by Goronszy (8) have demonstrated that when operating the batch reactor system with a continuous influent during all phases of the operational cycle, effluent quality is not seriously affected.

CANADIAN EXPERIENCE

The experience to be reported here is a first and concerns the retrofitting of a 9 m^3d^{-1} extended aeration plant designed to handle the flow from 5 residences at the Glenlea Agricultural Station just outside the city of Winnipeg (9). Erratic flows, infiltration resulting in solids washout, coupled with turbulent conditions in the intregal clarifier never permitted the plant to operate as originally envisaged. Mixed liquor suspended solids (MLSS) concentrations of 700 mgL^{-1} with effluent BOD_5 values of 90 mgL^{-1} were common (10).

In light of a 30/30 effluent requirement steps were taken in 1978 to convert the plant to operate in a batch reactor mode. Thirty days after conversion under manual control the MLSS had increased to 2 000 mgL^{-1}. The effluent BOD_5 was 22 mgL^{-1}. The plant was then retrofitted with automatic controls which resulted in being able to attain a MLSS concentration of 4 000 mgL^{-1} with an effluent BOD_5 of 15 mgL^{-1}. A year later (1980), routine examination showed effluent BOD_5 values of 7.5 mgL^{-1} and suspended solids of less than 5 mgL^{-1}.

Encouraged by these observations an extensive monitoring program for a number of batch reactor systems in Canada was supported by the Department of the Environment and the Department of Supply and Services. An intensive 90 day monitoring program at the Glenlea facility is one part of this program and is reported on as follows:

Flow

As shown in Figure 2, the daily average influent flow over the 90 day period to the reactor was 4.43 m^3. The flow variability was extremely high, varying from a low of 0.73 m^3d^{-1} to a high of 15.9 m^3d^{-1}. The flow ratios were:

low	: average	0.16
high	: average	3.59
high	: low	21.8

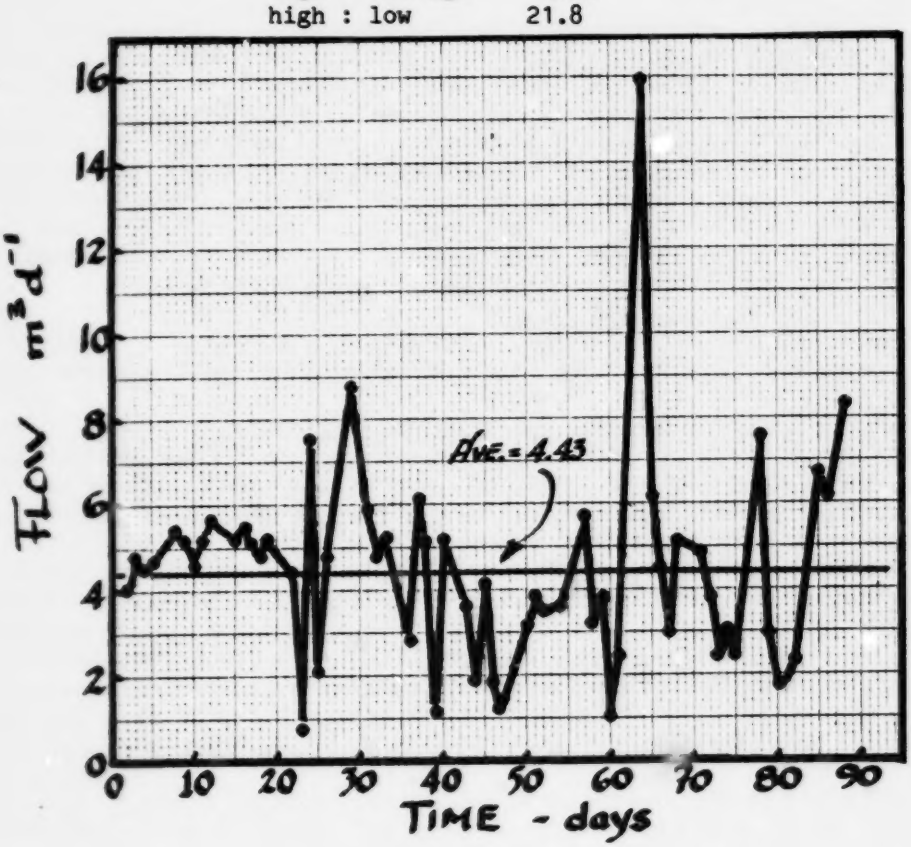

Fig. 2. Hydraulic load to the Glenlea plant.

The extremely high inflow of 15.9 m^3d^{-1} was caused by surface run-off infiltrating into a nearby manhole. The effluent BOD_5/SS values remained low at 4/4. During low influent rates the BOD_5/SS values were 7/1.

Biochemical Oxygen Demand (BOD_5)

Influent BOD_5 values varied from a low of 58 mgL^{-1} to a high of 800 mgL^{-1}, averaging 280 mgL^{-1} (Figure 3). This represent a relatively strong sewage which is typical of very small installations. On a mass basis the influent BOD_5 to the reactor averaged 1.14 kgd^{-1}. The effluent BOD_5 averaged 0.02 kgd^{-1} with concentrations over the 90 day testing period averaging 5 mgL^{-1}, never exceeding 10 mgL^{-1} but being as low as 2 mgL^{-1}.

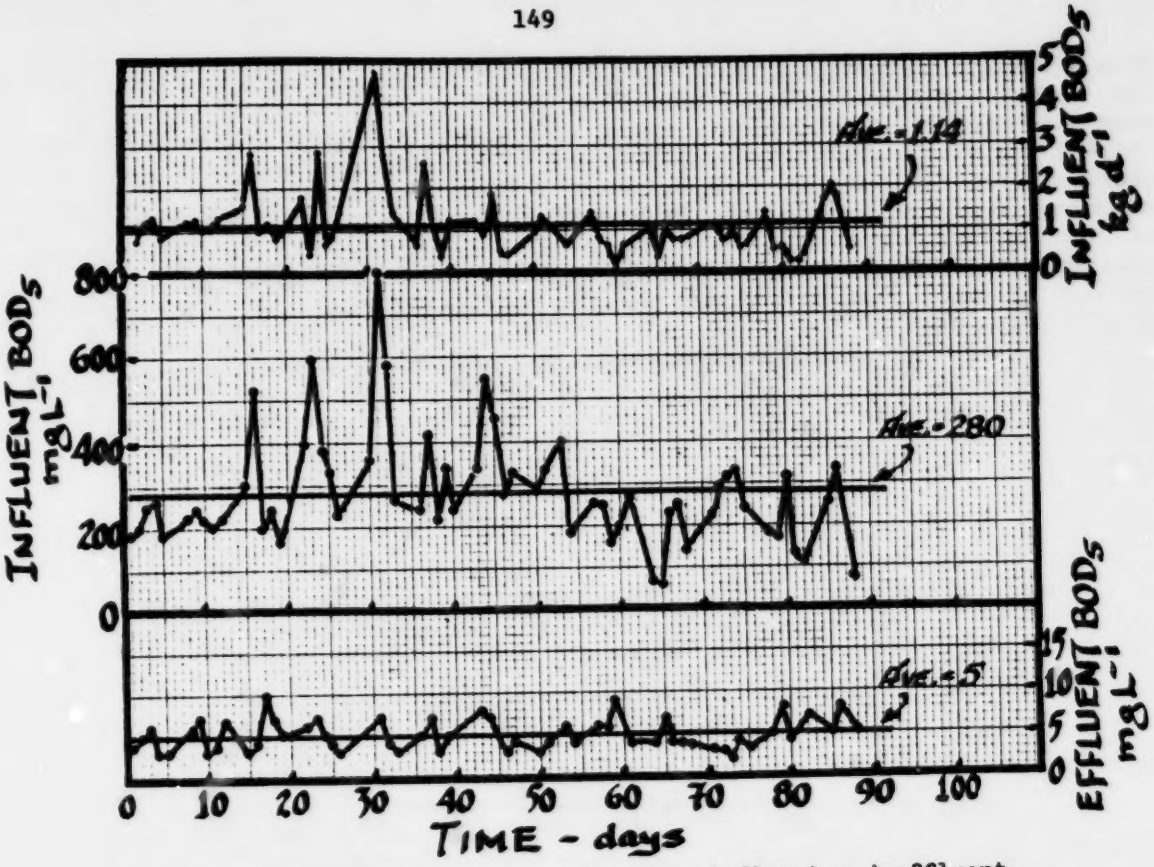

Fig. 3. BOD_5 concentrations and load - influent and effluent.

Process Loading

Conventionally, the food to microorganism ratio (F/M, kg BOD_5 applied/kg MLSS day) is used to define the organic loading to a biological process. Generally conventional activated sludge processes operate with an F/M equal to 0.5 whereas extended aeration processes are loaded at F/M = 0.05 and less. As shown in Figure 4 the initial F/M for the Glenlea plant was highly variable,

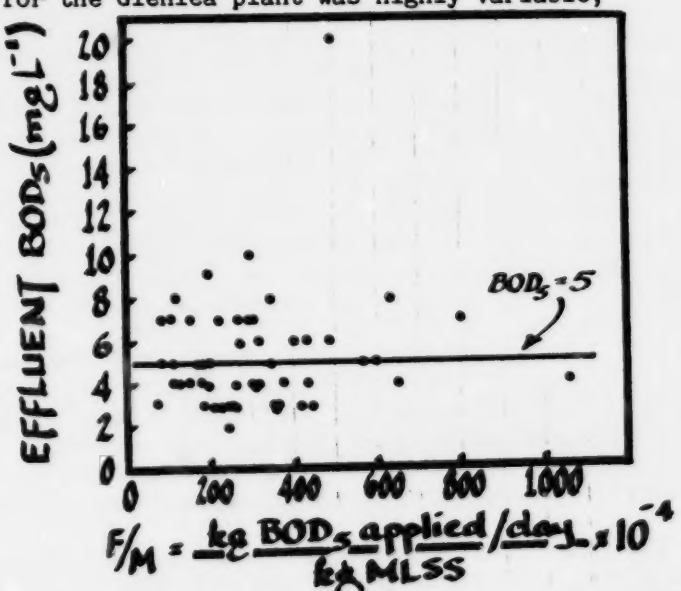

Fig. 4. Effluent BOD_5 as a function of F/M.

from a low of 0.006 to 0.105; a 17 fold variation. This of course is attributed to the highly variable influent BOD_5. In batch reactor technology the F/M ratio even under steady state organic loading conditions is dynamic and will decrease not only as the sludge mass increases from day to day but more importantly the F/M will decrease from its initial high value at fill to values of 0.0007 just prior to discharge when the organic load is exhausted. The process response over the range of loading conditions experienced is completely insensitive to influent organic loading either expressed as kg BOD_5 applied per day or grams of BOD_5 applied per cubic meter reactor volume per day. The effluent BOD_5 averaged 5 mgL^{-1} (Figure 5).

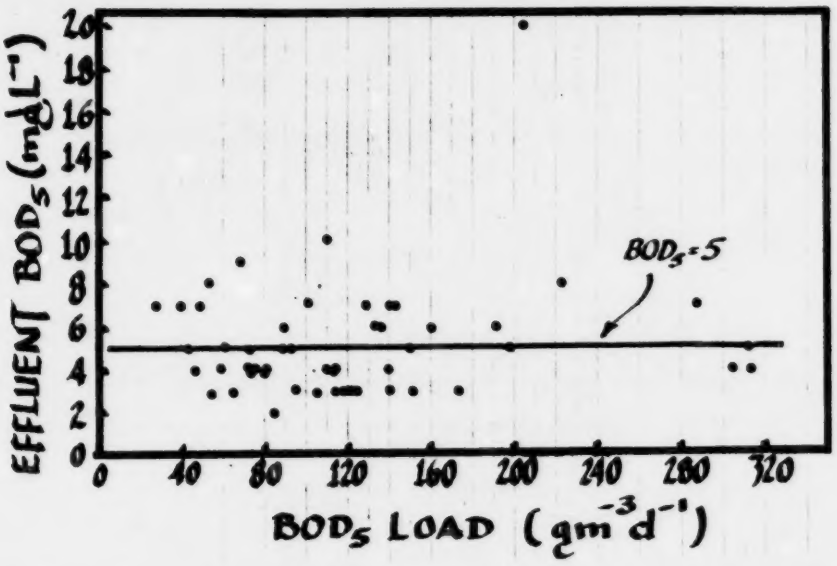

Fig. 5. Effluent BOD_5 as a function volumetric loading.

Sludge Growth

The amount of sludge accumulating in the system is of prime interest in that it establishes sludge wasting schedules. Figure 6 shows the increase in sludge concentration for 5 distinct periods over the 90 day monitoring period. As indicated by the data discontinuities sludge was wasted every 20 to 25 days. This was done with the objective of operating in a MLSS concentration range of 2500 - 4500 mgL^{-1}. The average MLSS concentration for the batch reactor can be taken as 3500 mgL^{-1}. The data are shown with 95% confidence bands and illustrate the initial problems encountered in trying to establish the value of the MLSS concentration. A fair degree of variability is evident. This is attributable to both sampling and analytical difficulties. As shown by the data in the later periods, this was overcome. Analysis of the the data show that the rate of solids production approximates on 1 kg for each kg of BOD_5 removed.

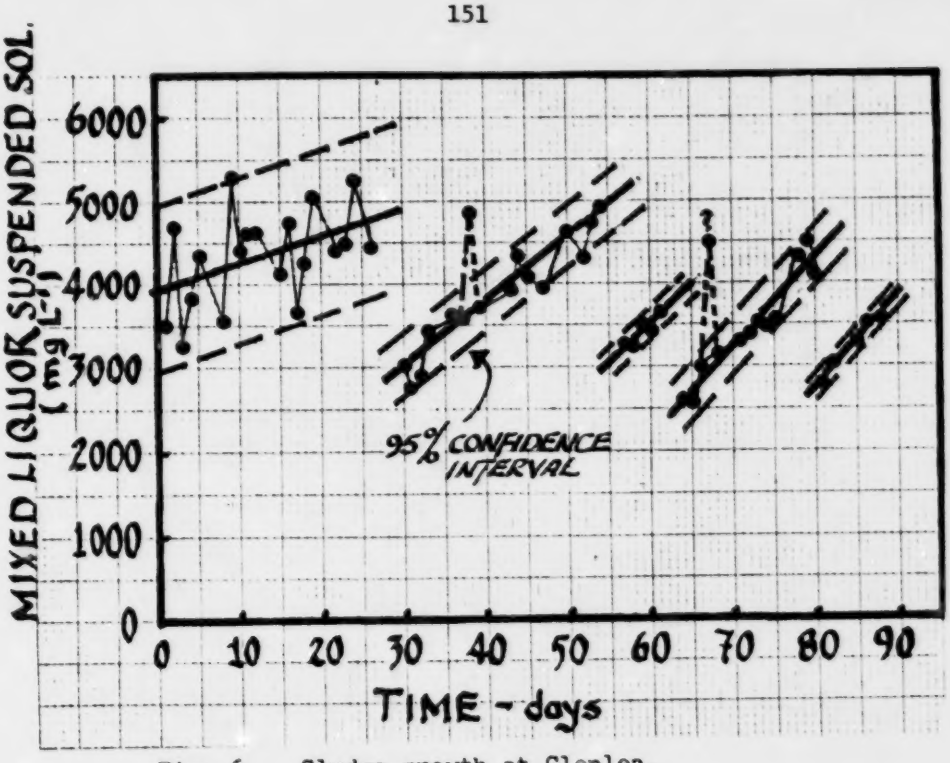

Fig. 6. Sludge growth at Glenlea.

Sludge Settleability

For suspended growth systems to operate properly, effective liquid/solid separation is essential. Thus, the microbial population must be maintained in a state where it has settling properties. A number of different methods are available to assess this settleability.

The method used here is the conventional sludge volume index (SVI) as determined by noting the volume occupied by 1 gram of sludge after settling for 30 minutes in a 1 liter cylinder.

$$SVI = \frac{SV_{30}\ mlL^{-1}}{SSi\ gL^{-1}} = \frac{ml}{g}$$

where:

SV_{30} = sludge volume after 30 minutes settling
SSi = initial sludge concentration

The sludge volume index is not a fixed index, but varies unpredictably with suspended solids concentration (11). Lee et al (12) recently examined and compared various quantitative techniques for assessing sludge settleability with the filament content of the sludge. They determined the total extended filament length (TEFL), which is a tedious operation, and proceeded to correlate it with the other methods of SVI determination. Their conclusion was that the best correlation between filament content and SVI was obtained when the diluted SVI was used. This means that the sludge was diluted to a concentration of 1.5 gL^{-1} ($SVI_{1.5}$).

Bearing the aforenoted in mind, especially that the SVI varies with solids concentration, the SVI data for the 90 day monitoring period at Glenlea is only an approximation. Nevertheless, the SVI, ranging from 70 to 160, attests to an excellent settling sludge. The average SVI was 116 (Figure 7).

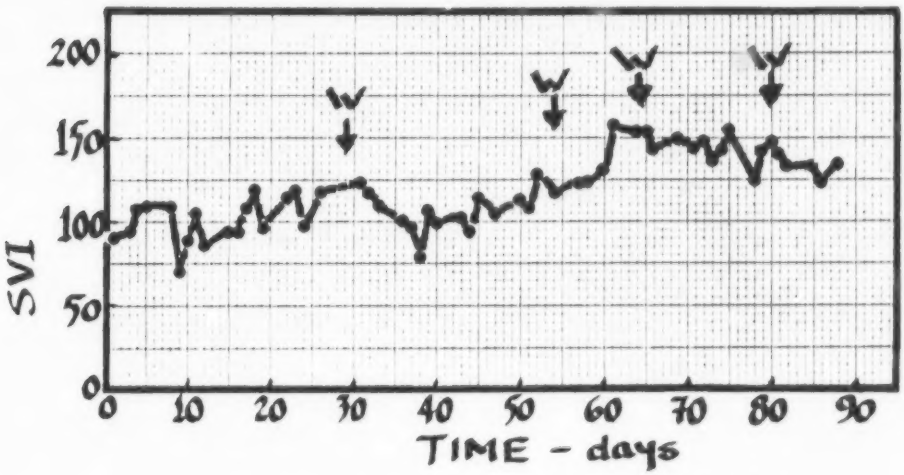

Fig. 7. Sludge volume index history at Glenlea.

Suspended Solids

As shown in Figure 8 influent suspended solids (SS) varied from 50 - 670 mgL^{-1}, averaging 177 mgL^{-1}. Effluent SS varied from 1 - 15 mgL^{-1} (1 value), and averaged 5.5 mgL^{-1}. This again attests to the excellent settling properties of the sludge.

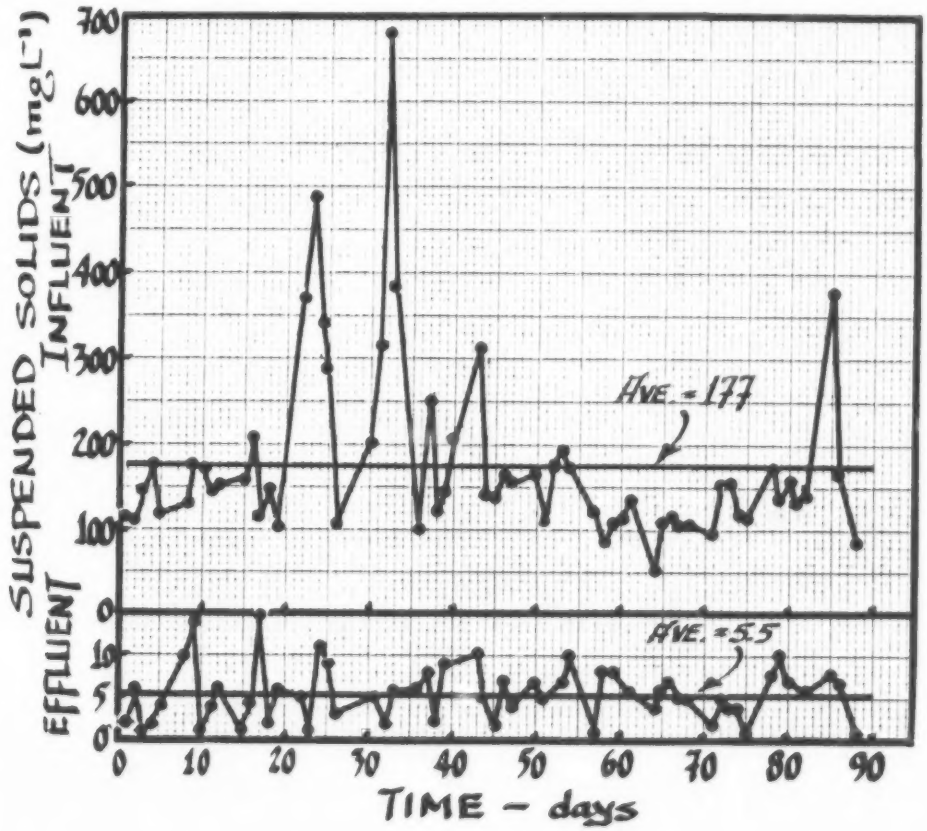

Fig. 8. Influent and effluent suspended solids.

Nitrification/Denitrification

It is well known that conditions conducive for growth of nitrifiers are:

Dissolved Oxygen	2 mgL^{-1}
pH	7 - 8.4

As well, sufficient alkalinity must be present, because theoretically 7 mg of alkalinity are used for each 1 mg of NH_3-N oxidized. For Glenlea alkalinity was not limiting to nitrification. The influent alkalinty varied from 100 mgL^{-1} to 600 mgL^{-1}, averaging 314 mgL^{-1}. The effluent alkalinity averaged 60 mgL^{-1} (Figure 9).

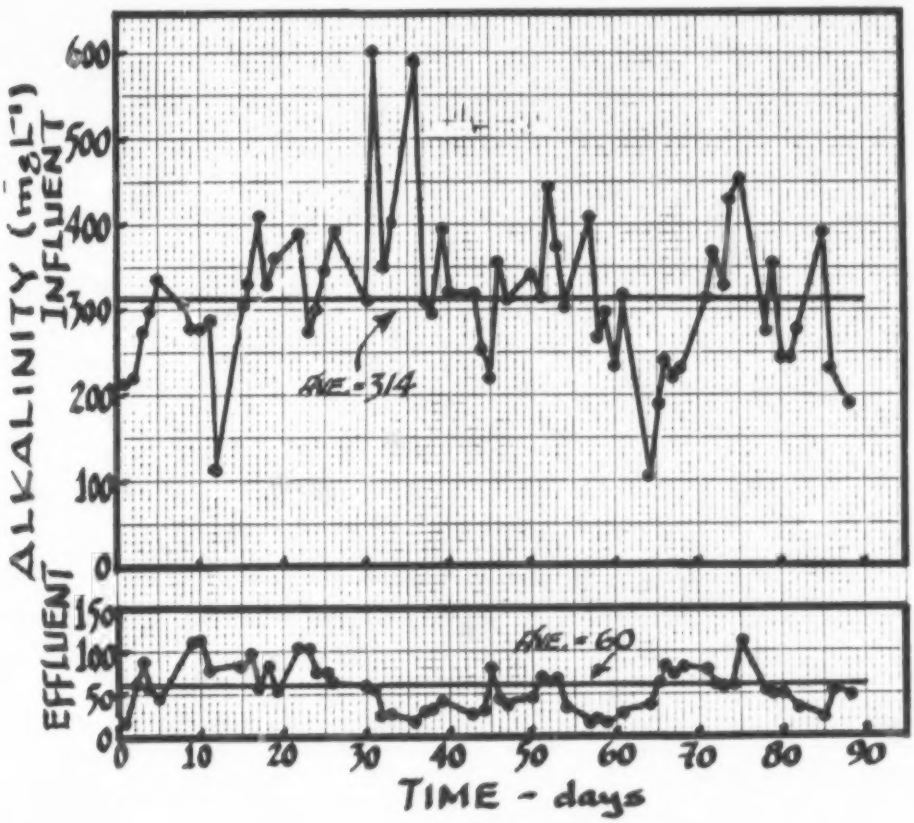

Fig. 9. Influent and effluent alkalinity.

As illustrated in Figure 10 wide fluctuations in influent ammonia concentration were experienced. The average concentration was 55 mgL^{-1}. This is an extremely strong sewage.

Effluent NH_3-N concentration varied between 1 and 9 mg/L (1 value), averaging 2.5 mgL^{-1} over the 90 day period.

Figure 11 shows the ratio of alkalinity removed to NH_3-N oxidized as averaging of 4.9 mg; as opposed to the theoretical value of 7 mg. This difference is attributed to denitrification since alkalinity will be partially restored by the destruction of nitrates.

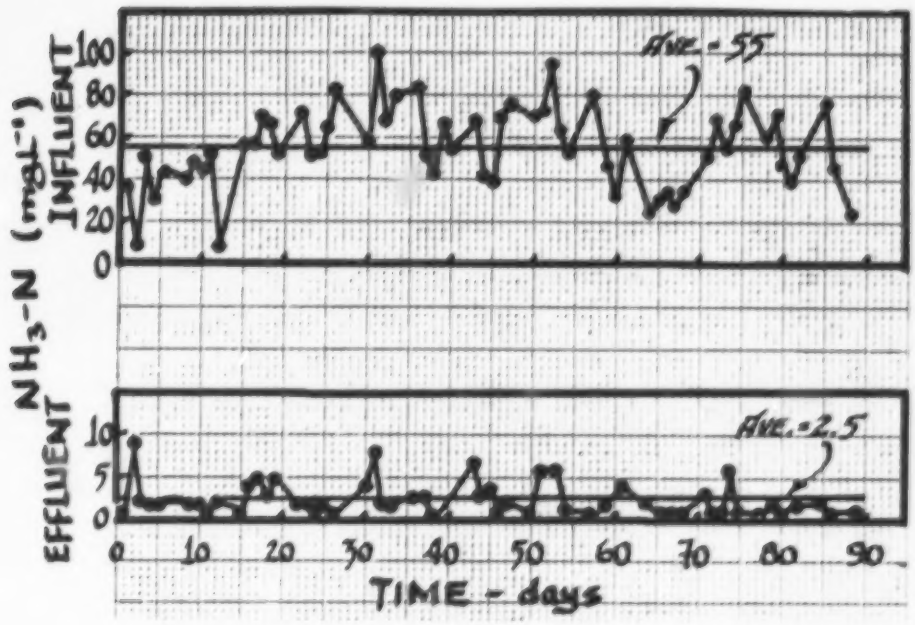

Fig. 10. Influent and effluent ammonia concentrations.

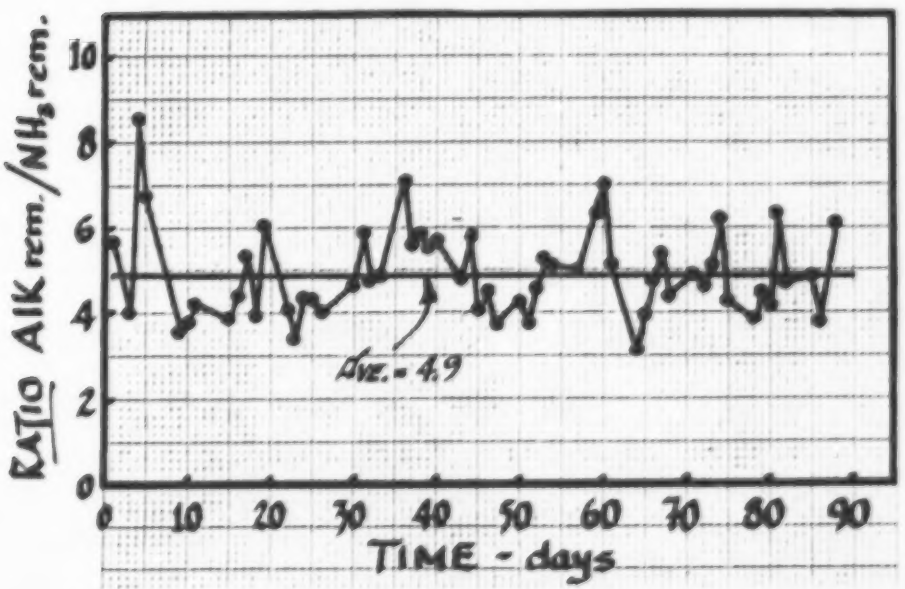

Fig. 11. Alkalinity removed to ammonia oxidized ratio.

A number of mathematical expressions for the growth of nitrifying organisms exist. One defines the net growth of nitrifiers under steady state conditions by:

$$\mu_m = 0.47\ e^{0.098\ (T-15)}$$

At 10°C, the net growth rate is calculated as 0.288 d^{-1}. The solids retention time (SRT), being the reciprocal of the growth rate, is 3.5 days. This then is the minimum required SRT for nitrification to occur.

At Glenlea, the liquid temperature varied from 9° - 11°C. SRTs were calculated to be 71, 84, 18 and 80 days. Clearly, conditions conducive to nitrification were present. Figure 12 shows the degree of nitrification. The average influent NH_3-N was 55 mgL^{-1}, the average NO_3-N formed was 31 mgL^{-1}. The influent pH of 7.8 was lowered to 6.5 due to hydrogen ion production during nitrification.

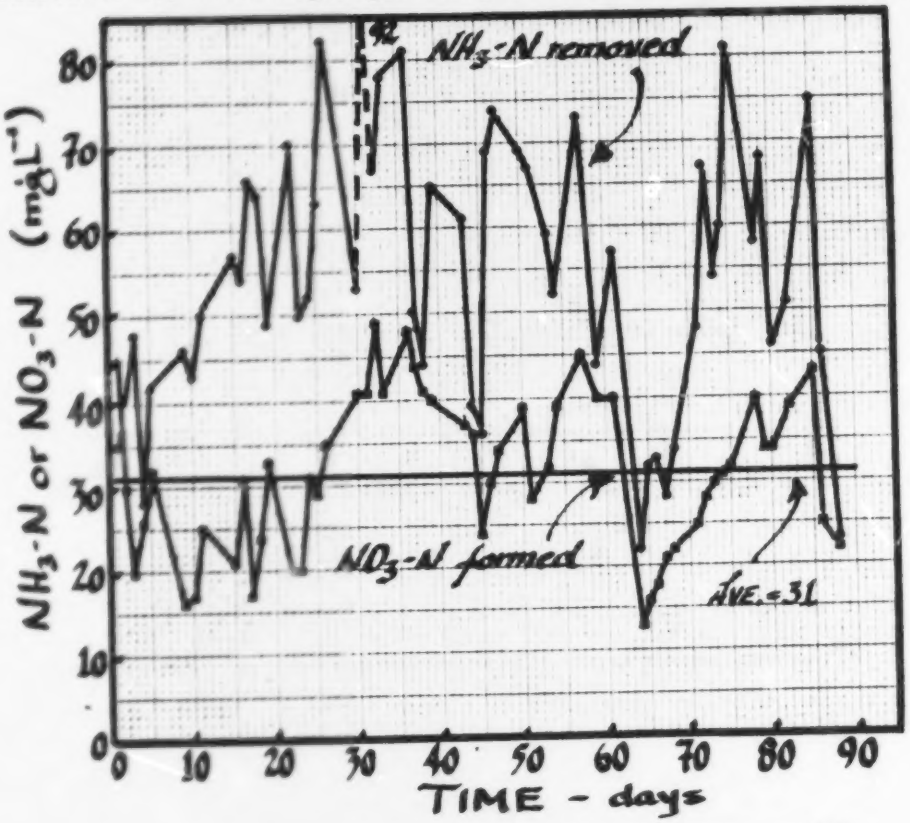

Fig. 12. Ammonia removed and nitrate formed profile.

As shown in Figure 13, a certain amount of denitrification also occured. This is illustrated by the data located to the right of the diagonal (more NH_3-N is removed than NO_3-N generated). As well, the effluent pH increased slightly to 6.6. It should be noted that no attempt was made at either optimizing nitrification or denitrification. The overall objective was to produce a 30/30 effluent. Clearly, full denitrification is possible and is only a matter of instituting the appropriate operational strategy. For example, a stir only cycle following the aeration cycle will denitrify the effluent. This again illustrates the desirability of striving for process flexibility by selecting equipment which will permit separation of aeration and mixing so that each may be controlled independently as required.

Phosphorus Removal

Quite unexpectantly, biological phosphorus removal exceeding 50% was observed. The influent total phosphorus (TP) averaged 11 mgL^{-1} and the effluent 5.4 mgL^{-1} (Figure 14). It can be calculated that approximately 3 mgL^{-1} of TP are required for cell synthesis. The fate of the remaining 2.6 mgL^{-1} TP is speculated to rest with 'luxury' uptake. As with denitrification, more controlled conditions coupled with a different operating strategy may further enhance phosphorus removal.

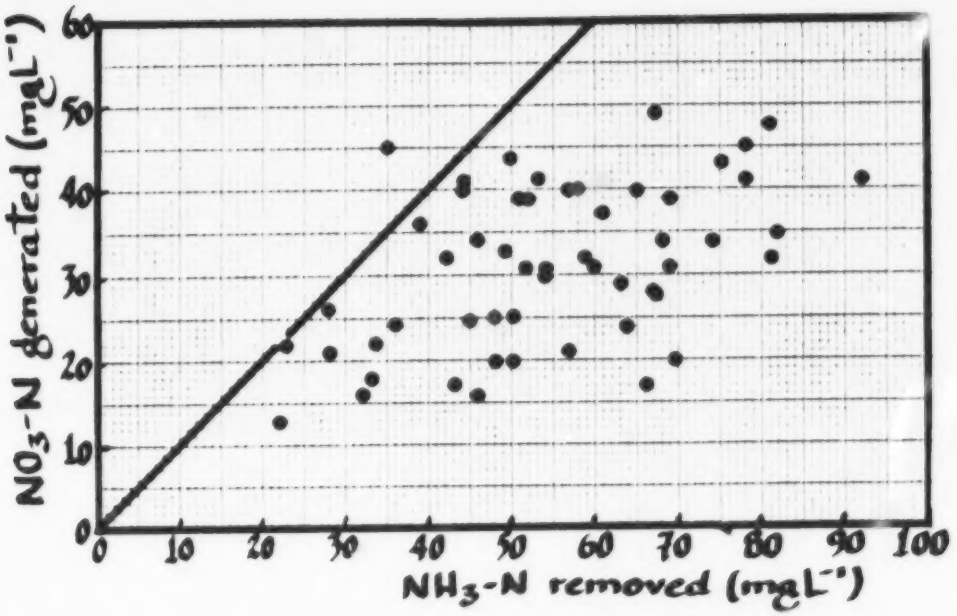

Fig. 13. Nitrate generated as a function of ammonia removed.

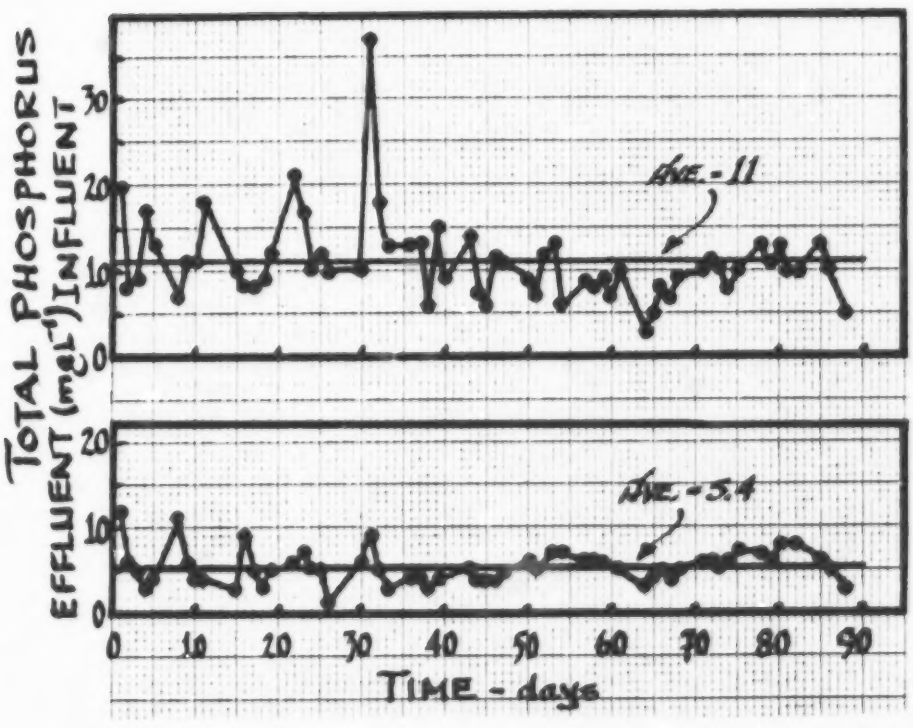

Fig. 14. Influent and effluent total phosphorus concentrations.

SUMMARY

The average process performance observed over the 90 day monitoring period is summarized in Table 1.

TABLE 1 Summary of Glenlea Monitoring Program

PARAMETER	UNITS	VALUE	
Flow	m^3d^{-1}	4.43	
MLSS	mgL^{-1}	3800	
SVI	mlg^{-1}	116	
F/M	kg BOD_5/ kg MLSS d	0.030	
		IN	OUT
BOD_5	mgL^{-1}	280	5
SS	mgL^{-1}	177	5.5
TKN	mgL^{-1}	77	14
NH_3-N	mgL^{-1}	55	2.5
NO_3-N	mgL^{-1}	-	31
TP	mgL^{-1}	11	5.4
pH	-	7.8	6.6

CONCLUSIONS

The long term performance observation of this batch reactor confirms that the biological treatment of waste originating from a small, high-strength, highly variable source poses no process problems. Indeed, a superior quality effluent as defined by BOD_5 = 5 mgL^{-1}, SS = 5.5 mgL^{-1} and NH_3-N = 2.5 mgL^{-1} was consistently achieved.

These observations confirm that the proper application of biological batch reactor systems is simple, inexpensive, requires a minimum of operator skill and attention, yet produces an effluent with characteristics normally associated with more complex biological waste treatment processes.

Application of this technology for process retrofit situations at existing plants is highly attractive when improved effluent quality is required. Batch reactors should be of particular interest to situations where seasonal, and/or high strength wastes from municipalities (small), resorts and industry require a quick, reliable, and cost-effective solution.

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ALKALINITY FEED FORWARD CONTROL OF ALUM ADDITION IN WASTEWATER TREATMENT

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INTRODUCTION

Phosphorus removal from wastewater is being practiced to control eutrophication of surface waters. Metal salt addition is the most widely used method of phosphorus removal in wastewater treatment, and alum is frequently selected as a precipitation chemical. In Norway there are more than 200 treatment plants based on chemical precipitation.

The objective of this paper is to present an alternative concept of alum addition based on alkalinity feedforward control.

BASIC CONSIDERATIONS

It is important to identify and quantify effects of the parameters that most directly establish required chemical dosage. The effect of final pH on the total phosphorus removal has been shown by several authors. Most of the data come from jar tests in which varying precipitant doses were followed by flocculation, settling, and often filtration. Literature reviews have been made (1) (2), and there seems to be a distinct connection between final pH and total phosphorus removal. Operating experiences in actual treatment plants show that a fairly narrow range of effluent pH values produces consistent and adequate effluent quality. Particle removal is even more sensitive to effluent pH than phosphorus precipitation (3). Thus the aim of the process control is to keep final pH value within the optimum range.

Chemical models for phosphorus precipitation have been validated by Ferguson and King (4), and by Balmér and Øren (5). Knowing the initial pH, alkalinity, and ortho phosphate, the required dosage to reach optimum final pH can be calculated by chemical equilibrium calculations.

Chemical models of the acid/base and precipitation reactions are used to show the relation between the dosage to reach pH 5.5 to 6.5 and the wastewater alkalinity (raw wastewater pH = 7.0 and phosphate concentration = 5.0 g P/m^3, Figure 1).

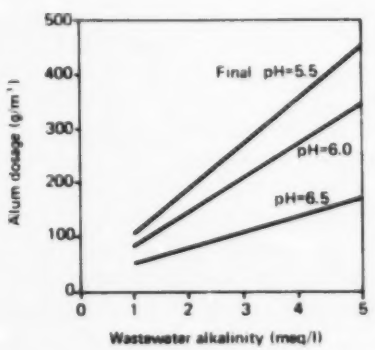

FIGURE 1. RELATION BETWEEN ALUM DOSAGE TO REACH DIFFERENT pH-VALUES AND WASTEWATER ALKALINITY FOR A TYPICAL MUNICIPAL WASTEWATER.

In practical situations there would not be perfectly straight line relations in Figure 1 because of differing pH and phosphate concentrations. However, it is apparent that wastewater alkalinity is the predominant variable affecting the alum dosage.

THE STATE OF ALUM ADDITION CONTROL

Variations in untreated wastewater flow and quality can be large. The alum requirement for phosphorus and suspended solids removal is often sensitive to these changes, and a process control system is required.

The control of precipitant addition may be divided into two groups,

- manual control - automatic control.

Manual control requires the operator to determine the status of the treatment or measure the quality of the effluent. Any deviations from desired operating conditions are manually corrected. Treatment plants with constant, clock programmed or flow proportional addition of chemicals fall into the group of manual control. These systems are often used in Norwegian treatment plants. If the water quality is monitored by sensors and the signals are used for automatic manipulation of the chemical dose, there is an automatic control system. An example of such a system is pH feedback control (closed loop control).

A survey of instrumentation and automation experiences in twelve chemical treatment plants in Norway has been carried out earlier. Ten plants were designed with pH feedback control and two with flow proportional dosage of chemicals. The actual situation showed, however, only three being operated with pH feedback control, seven with flow proportional, and at two plants chemicals were dosed at a constant rate with manual adjustments.

The most frequent reason for changing the control systems was trouble with the pH sensor. Another reason may be insufficient operator training and operation manuals on how to maintain and tune closed loop control. Such systems require more skilled operation and maintenance and good understanding of the principles of pH measurement.

ALKALINITY FEEDFORWARD CONTROL

The investigations reported above document the widespread abandonment of pH feedback control system and poor performance of some still in use. There was an obvious need to develop alternative control methods for chemical addition. One such system, flow proportional with feedforward dosage based on alkalinity, is an open loop control which avoids most of the problems of oscillations between the sensor and the chemical dosage with the resulting oscillations in effluent quality. Calibration of open loop systems is inherently simpler. For practical purposes it is apparent that wastewater alkalinity is the predominant parameter affecting chemical dosage, and the correlation between wastewater alkalinity and alum dosage is almost linear.

A control scheme based on alkalinity feedforward control is proposed (Figure 2). This control method requires an alkalinity measuring device. Since the measuring device shall be applied directly on pre-settled sewage, the reliability is of great importance. Automatic measuring units for wastewater are still not available and therefore alkalinity is measured indirectly as conductivity.

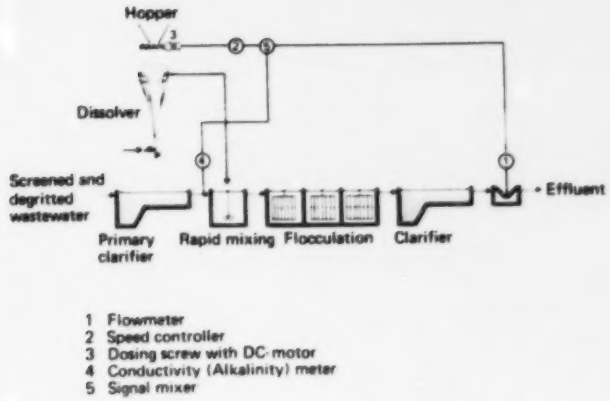

FIGURE 2. SCHEME OF AN ALKALINITY FEEDFORWARD CONTROL SYSTEM.

FULL SCALE INVESTIGATION

Initial Improvements

A technical assistance project has been carried out at Sørumsand treatment plant, a primary-chemical treatment plant designed for 6,000 persons. Granulated alum is dosed flow proportionally via an upflow dissolver. Experiences from the treatment plant have shown very unstable alum dosage and as a result, oscillating pH and bad process reliability. Deviations from desired operating conditions were manually corrected by changing the proportionality factor several times a day.

The main deficiencies and proposed improvements are shown in the following table.

Problem	Proposed improvements
- Too large measuring range of flow meter resulted in unreliable measurements during dry weather conditions.	- Reduce measuring range by replacing the measuring profile.
- Too large operating range of alum feeder caused unlinear performance at low flow rates.	- Reduce operating range by altering gear conversion ratio.
- Unsatisfactory performance of flow proportional dosage due to large alkalinity variations.	- Change to alkalinity feedforward control.

Conductivity was measured by a 4-electrode measuring device, range 0-1,000 µS/cm, with automatic temperature compensation. Signals (0-20 mA) from the bulk flow meter and the conductivity measuring unit were mixed in a signal mixer, and chemical dosage was controlled by the resulting signal.

The proposed improvements have been undertaken involving the following capital costs: (One USD = 7.4 NOK.)

Conductivity measuring system	NOK	10,000
Signal mixer	NOK	2,000
Installation	NOK	2,000
Sum capital costs	NOK	14,000

The design improvements and the automatic control system (Figure 2) have been reported to give a remarkably better process efficiency at lower operating costs.

Investigation Programme

The efforts were followed up by an investigation concerning control performance, process efficiency, and operating experiences at Sørumsand treatment plant. Experiments were carried out during three periods, each lasting two days, including two snow melting periods and one dry weather period. The treatment plant performance is controlled by a routine programme.

Results

Wastewater alkalinity and specific conductivity were measured simultaneously during the experiments (Figure 3).

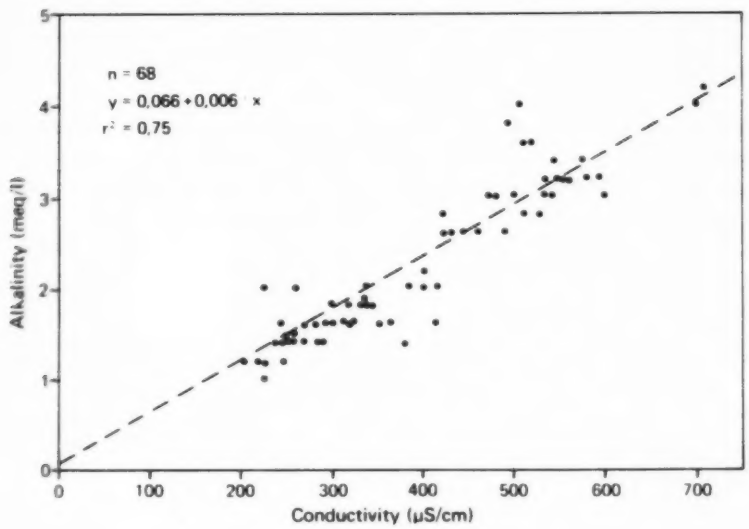

FIGURE 3. RELATION BETWEEN WASTEWATER ALKALINITY AND SPECIFIC CONDUCTIVITY FOR SØRUMSAND TREATMENT PLANT.

The data points do not fall on a perfect straight line, however, it was found to be suitable for control purposes. Due to the dilution effect inlet alkalinity decreases by increasing flowrates, as shown in Figure 4.

Regarding the process control alum dosage is varying with flowrate as shown in Figure 5. The alum dosage varied between 60 and 300 g/m^3. The resulting pH-value is nearly constant (6.4-6.6) over the flow range, and this indicates the efficiency of the control system.

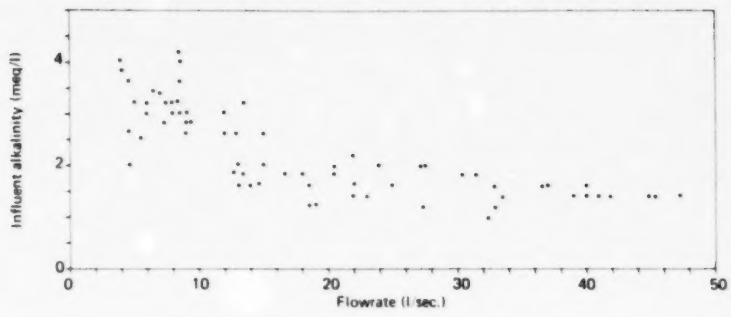

FIGURE 4. INFLUENT ALKALINITY vs. FLOWRATE.

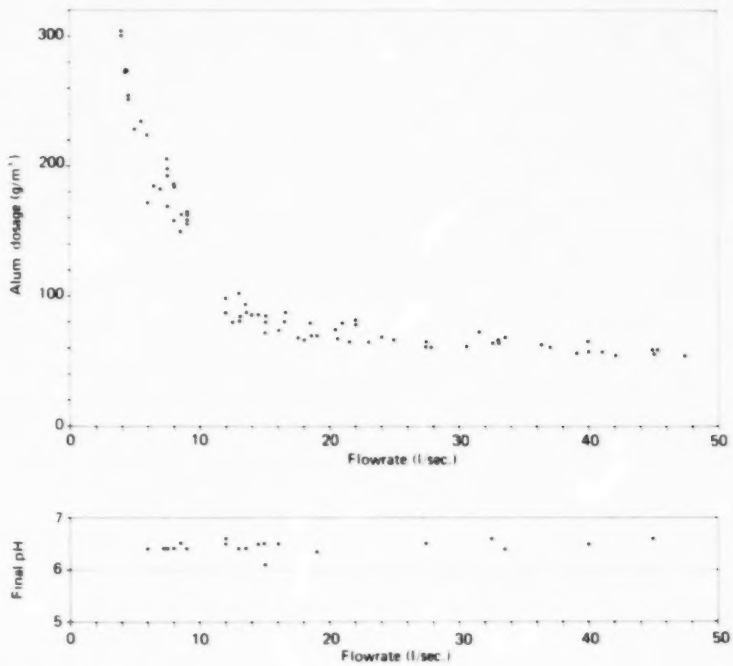

FIGURE 5. ALUM DOSAGE AND FINAL pH vs. FLOWRATE DURING THE TEST PERIODS.

The long term improvements according to the routine programme show an overall increase of total phosphorus removal efficiency from 96 to 99 percent. Frequency plots of outlet ortho phosphate concentration and Secchi-depth in final clarifier are shown in Figure 6.

By changing to automatic feedforward control the average alum consumption was reduced from 270 g/m^3 to 190 g/m^3. Annual amount of wastewater treated is about 200,000 m^3. The new control system is now permanently installed at Sørumsand treatment plant, and according to the operators the need for manual corrections is negible. At present other Norwegian treatment plants are also changing to alkalinity (conductivity) feedforward control.

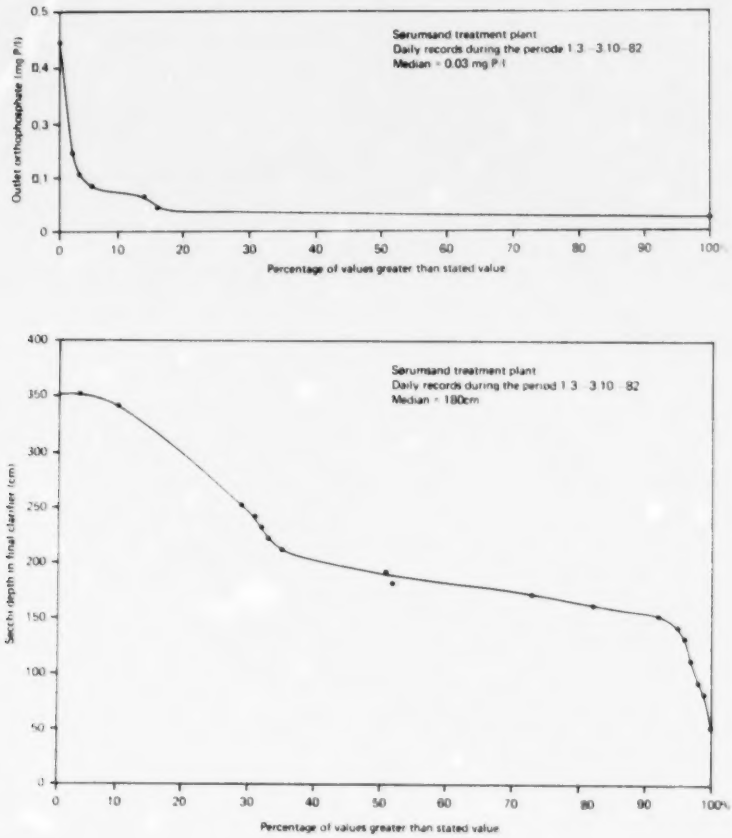

FIGURE 6. FREQUENCY PLOTS OF EFFLUENT ORTHO PHOSPHATE AND SECCHI DEPTH IN FINAL CLARIFIER.

CONCLUSIONS

The conclusions of the project were:

- The accuracy of the flowmeter and dosage unit is of basic importance in establishing a reliable control system.
- Alkalinity feedforward control resulted in significantly better process efficiency and reliability compared to flow proportional dosage.
- Due to the lack of automatic alkalinity measuring devices for wastewater, conductivity should be considered as control parameter. This requires a good correlation between conductivity and alkalinity.

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LA DEPHOSPHATATION BIOLOGIQUE

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INTRODUCTION

Les principaux objectifs à atteindre dans le domaine du traitement des eaux résiduaires urbaines ont toujours été l'élimination de matière en suspension et de la matière organique biodégradable, DBO. Les installations d'épuration classiques, consistant en un traitement primaire (mécanique) et secondaire (biologique) permettent d'atteindre ces buts sans trop de problèmes. Cependant, il s'avère de plus en plus souvent nécessaire d'éliminer aussi l'azote et le phosphore, afin d'éviter la détérioration de la qualité des eaux de surface par des phénomènes d'eutrophisation.

L'azote peut être éliminé biologiquement par le procédé de nitrification et dénitrification, dans lequel l'azote s'échappe finalement sous forme gazeuse (N_2). Il ne semble au contraire pas possible de transformer le phosphore dans une forme volatile par traitement biologique. L'élimination du phosphore devra donc être effectuée par inclusion dans une phase solide, par exemple dans une boue. La précipitation physico-chimique par ajout de chaux, de sels de fer ou d'aluminium, parait actuellement la technique la plus évidente. La production de boues supplémentaires et les coûts d'achat des réactifs nécessaires sont cependant des inconvénients considérables liés à cette technique. Il semble donc intéressant d'examiner plus profondément les possiblités d'obtenir une boue biologique enrichie en phosphore, par modification du procédé des boues activées sans ajoût de réactifs chimiques. Une meilleure compréhension des mécanismes intervenants permettra de mieux maîtriser la mise en oeuvre et le fonctionnement de cette "déphosphatation biologique".

DEVELOPPEMENT DES PROCEDES EXISTANTS

Pendant la deuxième moitié des années 60, des rendements de déphosphatation extraordinaires ont été mentionnés pour quelques stations d'épuration aux Etats-Unis (Vacker, Scalf...). Il s'agissait chaque fois de stations à forte charge, non nitrifiantes, avec un bassin d'aération en flux piston. Comme l'aération consistait en une insufflation d'air et que le taux de recyclage était faible, il y avait toujours à l'entrée du bassin d'aération une zone où la concentration en oxygène dissous était presque nulle. Il était constaté une augmentation de la concentration en phosphore dans cette zone, suivie par une diminution très rapide dans la partie du bassin d'aération où la concentration en oxygène dissous était plus élevée. Des concentrations finales en phosphore inférieures à 1 mg/l étaient ainsi obtenues, sans qu'il fut cependant possible d'expliquer ce phénomène particulier.

A partir de ces observations, plusieurs procédés ont été développés. Ils font appel aux réactions biochimiques pour obtenir une surélimination du phosphore.

DIFFERENTS PROCEDES DE MISE EN OEUVRE ENVISAGEABLES

Procédé "Phostrip" (Fig. 1)

Dans le procédé "Phostrip" les boues décantées dans le clarificateur final sont envoyées dans un épaississeur gravitaire.

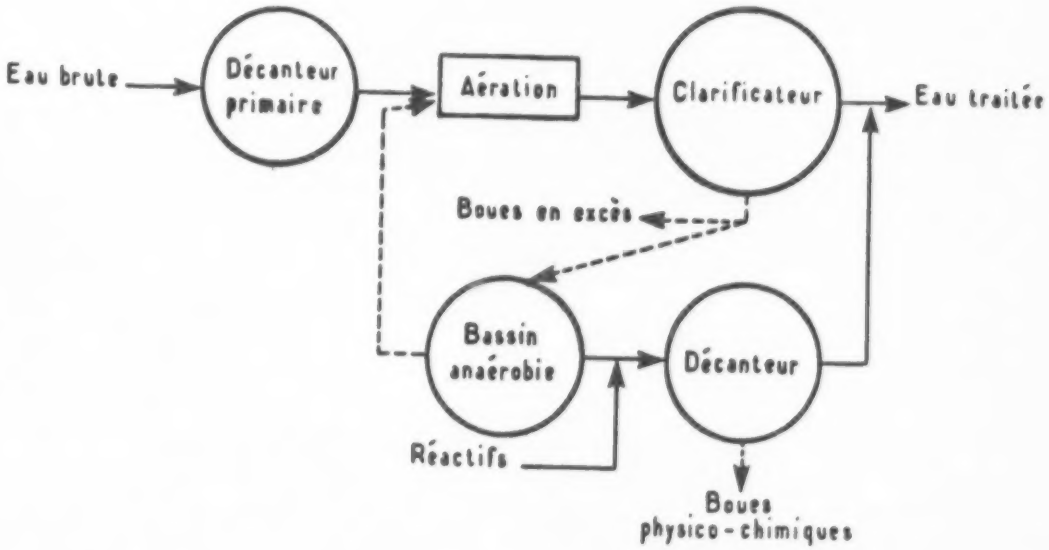

Fig. 1 : Procédé "Phostrip"

Les conditions anaérobies dans cet épaississeur provoquent un relargage de phosphore dans le milieu extracellulaire. Le surnageant de l'épaississeur est traité à la chaux afin de précipiter les phosphates, puis recyclé en tête de la station ou rejeté avec l'effluent traité. Les boues épaissies, qui se sont partiellement vidées de leurs réserves de phosphore intracellulaires, sont recyclées dans le bassin aéré, où elles absorbent le phosphore contenu dans l'eau brute. Le temps de séjour de la boue dans le "stripper" anaérobie est de 10 à 24 heures. L'avantage de ce procédé par rapport aux traitements classiques se situe dans le fait que la quantité de chaux nécessaire pour la précipitation est plus faible : cette quantité est en effet fonction plutôt du pouvoir tampon de l'alcalinité de l'eau brute que de la quantité de phosphates à précipiter. Comme le débit d'eau sur lequel la précipicipitation est faite n'est que 10 à 20 % du débit d'eau brute, des économies importantes peuvent être faites sur les doses d'agent précipitant nécessaires.

Procédé A/O (Fig. 2)

Le procédé A/O (anaérobie/oxie) comporte simplement une zone anaérobie où l'eau brute est mélangée avec les boues recyclées et une zone aérée après.

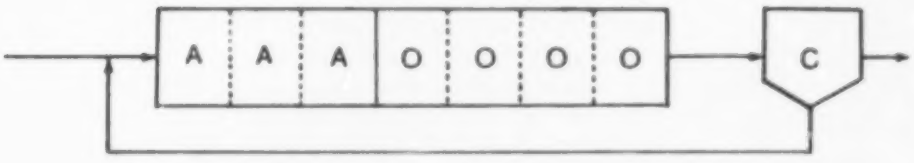

Fig. 2 : Procédé A/O

A : Zone d'anaérobiose
O : Zone aérée
C : Clarificateur

Les deux zones sont divisées en plusieurs compartiments, simulant ainsi l'hydraulique d'un flux piston. La zone anaérobie est couverte afin de limiter l'apport d'oxygène par le contact avec l'air. Le relargage du phosphore dans ce système est plus rapide que dans le procédé "Phostrip": un temps de séjour en anaérobiose de 30 minutes à une heure est conseillé pour le dimensionnement. Cette différence est due à l'apport de matière organique dans le bassin anaérobie du procédé A/O : les résultats d'essais mentionnés plus loin dans cet article ont démontré l'effet stimulateur de l'apport d'un substrat carboné sur le relargage du phosphore en anaérobiose. Dans le bassin aéré, les phosphates sont réabsorbés. Des effluents traités contenant moins de 1 mg P/l ont été obtenus (Barth 1982). Les boues en excès produites contiennent environ 7 % de phosphore par rapport aux matières sèches, tandis que la teneur normale se situe entre 2 et 3 %.

Procédé "Bardenpho" (Fig. 3)

En 1974, il était rapporté que des stations du type "Bardenpho" étaient particulièrement performantes quant à l'élimination du phosphore. Ce procédé

était conçu en principe pour nitrifier et dénitrifier, et consiste en 4 bassins en succession, dont seulement le deuxième et le quatrième sont aérés.

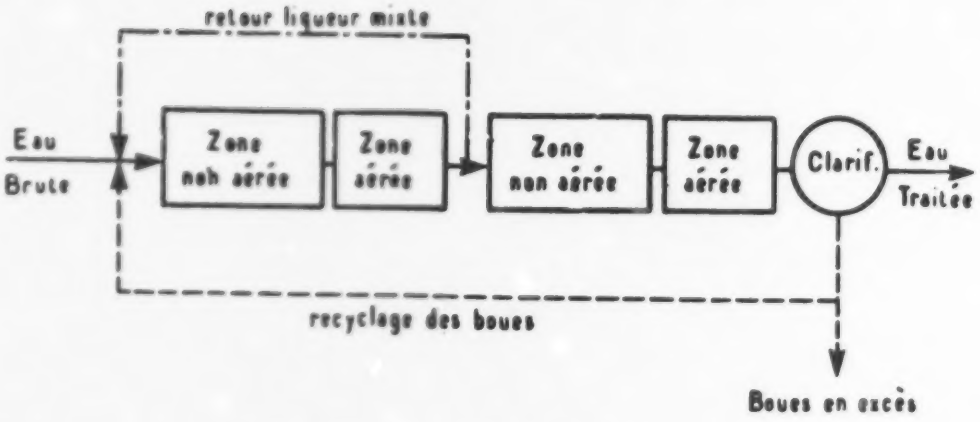

Fig. 3 : Procédé "Bardenpho"

La liqueur mixte sortant du deuxième bassin est partiellement recyclée dans le premier bassin (dit bassin d'anoxie) où les nitrates formés sont dénitrifiés à l'aide de la matière organique présente dans l'eau résiduaire. Le deuxième bassin anoxique sert à dénitrifier le reste de la liqueur mixte, cette fois par dénitrification endogène, utilisant la matière cellulaire elle-même comme substrat organique. Le dernier bassin réarère la boue avant décantation. Avec ce procédé, une élimination quasi-totale de l'azote est possible. Il a été trouvé, comme profil des concentrations en phosphore dans les bassins consécutifs : 2,5 ; 2,5 ; 30 et 0,3 mg P/l (Barnard (1974).
Ce serait donc surtout le troisième bassin qui ferait office de zone anaérobie, bien qu'il n'y ait presque plus de matière organique en solution. La plupart des chercheurs travaillant sur ce système n'ont d'ailleurs pas réussi à obtenir un relargage de phosphore dans ce bassin anoxique secondaire, même en absence de nitrates : une bonne dénitrification est une condition nécessaire mais pas suffisante pour obtenir une déphosphatation accrue dans le système "Bardenpho". La faible fiabilité du système sur le plan de l'élimination du phosphore est due à ces problèmes avec les nitrates.

Procédé "Phoredox" (Fig. 4)

Pour éviter ces difficultés, le procédé "Phoredox" inclut un bassin anaérobie en tête d'une station "Bardenpho" où l'eau résiduaire est mise en contact avec les boues recyclées du clarificateur.

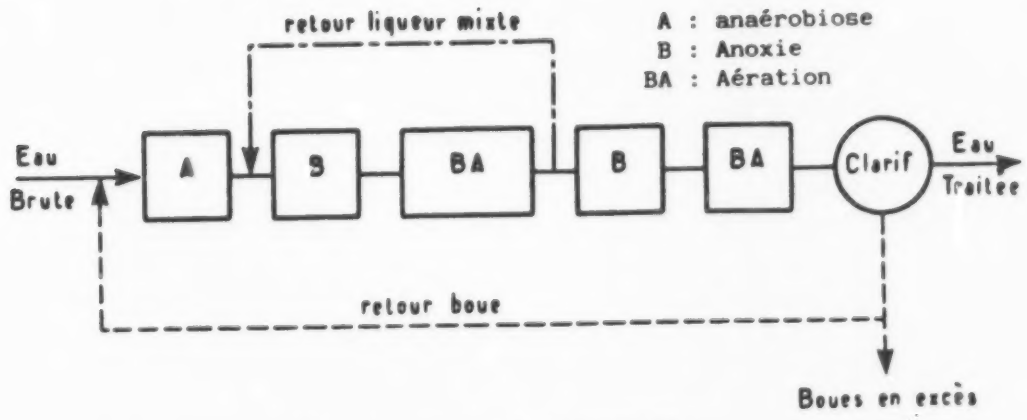

Fig. 4 : Procédé "Phoredox"

L'apport de nitrates étant ainsi limité, les conditions anaérobies sont mieux gardées que dans les bassins anoxiques du "Bardenpho" original. Un temps de séjour de 1 à 2 heures dans le bassin anaérobie (appelé aussi bassin de fermentation) est recommandé (Barth 1982). Dans une version simplifiée du procédé "Phoredox", les bassins de dénitrification endogène et de réaération sont supprimés, ce qui réduit considérablement les coûts d'investissement (fig.5). La dénitrification dans le deuxième bassin anoxique du "Bardenpho" étant dix fois plus lente que dans le premier bassin, une légère augmentation du volume de ce bassin anoxique primaire peut avantageusement compenser la capacité de dénitrification perdue par omission du bassin anoxique secondaire.

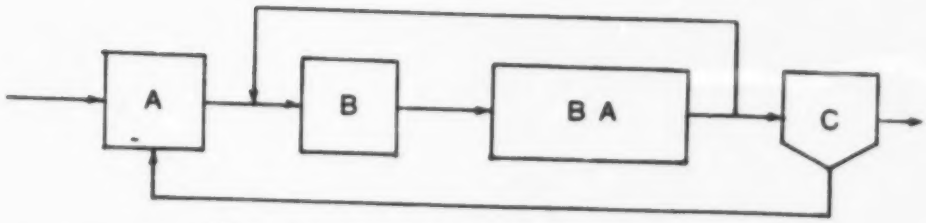

Fig. 5 : Procédé "Phostrip modifié"

A : Zone anaérobie
B : Zone anoxie
BA : Zone aérée
C : Clarificateur

Procédé "UCT" (Fig. 6)

Même avec un bassin de fermentation à l'entrée de la station, un bon rendement de déphosphatation ne peut pas toujours être assuré dans les procédés "Phoredox et Phoredox modifié". En effet, si la dénitrification ne réussit pas à procurer un effluent traité titrant très faible en nitrates, il y a possibilité de perturbation des conditions anaérobies dans la zone de fermentation par les nitrates recyclés avec les boues décantées. A base de considérations théoriques, une équipe sudafricaine a calculé un rapport limite NTK/DCO de 0,08, en dessus duquel une dénitrification complète dans un procédé "Phoredox ou Bardenpho" n'est plus possible (Marais 1982). Le rapport NTK/DCO habituel est de 0,06 à 0,08 pour une eau urbaine non décantée, et de 0,09 à 0,11 après décantation primaire. Une bonne dénitrification, nécessaire pour une bonne déphosphatation, sera donc souvent difficile, voire impossible, donnant ainsi lieu à des performances déphosphatantes aléatoires.

Cette influence négative des nitrates sur le rendement d'élimination du phosphore pourrait être évitée en mélangeant l'eau brute dans le bassin anaérobie non pas avec les boues reculées du clarificateur, mais avec la liqueur mixte dénitrifiée sortant du bassin d'anoxie (Procédé UCT : University of Cape Town).

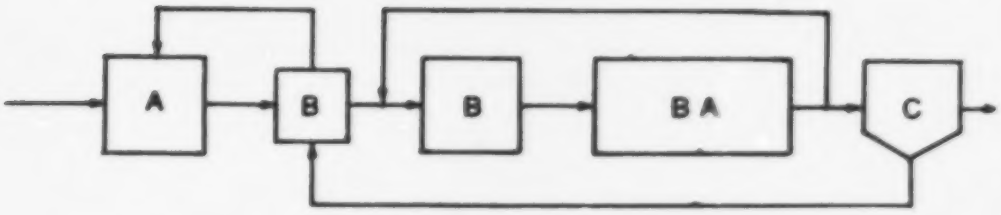

Fig. 6 : Procédé "UCT"

A : Zone anaérobie
B : Zone anoxie
BA : Zone aérée
C : Clarificateur

La bassin d'anoxie reçoit alors la liqueur mixte recyclée du bassin d'aération, le recyclage des boues décantées et l'effluent bassin anérobie. Théoriquement ce système pourrait fonctionner correctement pour des rapports NTK/DCO jusqu'à 0,14, en éliminant tous les nitrates dans la liqueur qui va être recyclée en anérobiose.

ETUDE AU LABORATOIRE

Pilote

Afin d'examiner certains aspects cinétiques et mécaniques du phénomène de la surélimination biologique du phosphore, des essais ont été effectués à l'échelle laboratoire sur un pilote du type "Phoredox modifié" (fig.7). Le pilote comporte deux bassins non aérés de 8 l servant de bassins d'anaérobiose et d'anoxie. Le bassin d'aération est divisé en quatre compartiments, afin de simuler un régime de flux piston. Le volume du bassin aéré est de 12 l. Le pilote est alimenté à 2 l/h avec une eau résiduaire urbaine, parfois supplémentée avec du lait écrémé pour maintenir une DCO suffisante. La composition moyenne de l'eau brute est donnée dans le tableau 1.

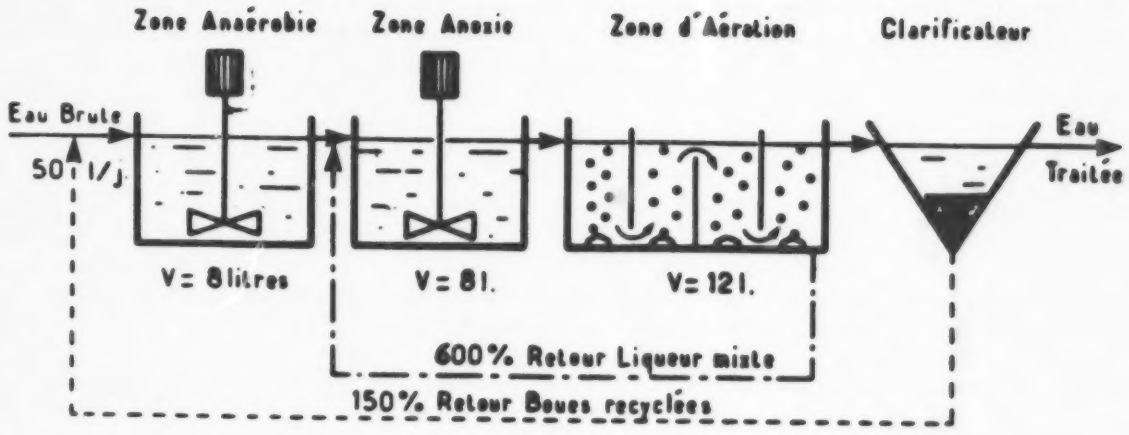

Fig. 7 : Schéma du pilote

55 % de la DCO de l'eau brute est soluble, et la DBO_5 représente 44 % de de la DCO. Le rapport NTK/DCO est de 0,09, ce qui est légèrement supérieur à la valeur de 0,08 qui devrait selon Marais (1982) garantir un bon fonctionnement.

Tableau 1 - Résultats obtenus sur pilote

Paramètres (mg/l)	Eau brute	Eau traitée	% élimination
DCO	670	48	93
MES	210	15	93
P total	12,6	1,85	85
NTK	60	2,35	96
$N\text{-}NO_3$	0	3,4	-

Cependant, des résultats consistants et encourageants ont été obtenus : le tableau 1 donne des rendements d'épuration moyens pour une période de

six semaines de fonctionnement stable. Hors de l'élimination excellente de la DCO et des matières en suspension, il est constaté une élimination de 85 % du phosphore et de 91 % de l'azote total. Pendant cette période, la concentration en matières séches dans le pilote était 6 g/l, à 77 % de matière volatile. La production de boues était en moyenne 9,19 g/j, correspondant à un âge de boues aérées de 7,8 jours. Au vue des teneurs en matières en suspension des eaux résiduaires et de l'élimination de la DBO5, cette production de boues correspond théoriquement à une charge massique de 0,19 kg DBO/kg MS.j (documentation Société Degrémont). La charge massique calculée sur le pilote est de :

$$\frac{48\ l/j \times 670\ mg\ DCO/l \times \frac{0,44\ mg\ DBO5}{mg\ DCO}}{12\ l \times 6\ g\ MS/l} = 0,196\ \frac{kg\ DBO}{kg\ MS.j}$$

si on ne tient compte que de la boue aérée dans ce calcul. La production de boues en excès dans une installation de déphosphatation biologique n'est donc pas supérieure à celle d'une station classique fonctionnant à un même âge de boues aérées.

Aucun ensemencement spécifique n'a été nécessaire pour obtenir une déphosphatation accrue : le pilote a été démarré avec une boue provenant d'une station à boues activées classique, fonctionnant à faible charge. Dès les premiers jours d'opération, le rendement d'élimination du phosphore s'est graduellement amélioré, pour arriver à des concentrations inférieures à 1 mg/l après une douzaine de jours. Il est donc permis de dire que les organismes responsables de la déphosphatation sont déjà naturellement présents dans une boue activée ou dans l'eau résiduaire urbaine, et qu'ils deviennent plus actifs ou plus nombreux sous l'influence des conditions anaérobies/aérobies alternantes. Le même phénomène pourrait être déduit de l'évolution du profil des concentrations en phosphore dans les différents bassins (fig.8) : les premiers jours d'opération, le relargage du phosphore en anaérobiose est plutôt faible, ainsi que l'élimination globale du phosphore. Au cours du temps, la quantité de phosphore relarguée devient plus importante, mais la réabsorption derrière est devenue plus rapide aussi, permettant une élimination presque totale du phosphore. La relation étroite entre le relargage du phosphore dans la zone anaérobie et l'élimination globale sur l'installation est illustrée aussi dans la figure 9 qui donne la quantité de phosphore relarguée (calculée comme la différence des concentrations en phosphore à l'entrée et à la sortie du bassin anaérobie) et celle éliminée, pour une période de fonctionnement aléatoire (dû à un rapport NTK/DCO trop élevé). On voit clairement le décalage de 1 à 2 jours qui existe entre le relargage et l'élimination du phosphore : le relargage au jour X ne provoque une déphosphatation accrue que les jours X + 1 ou X + 2. On pourrait dire que le relargage du phosphore induit un stockage de "capacité de déphosphatation" dans la boue. Cette "capacité" est consommée ultérieurement donnant lieu à une surélimination du phosphore. L'allégation de Barnard (1976) disant que, en amont de l'aération finale, une étape anaérobie doit être incluse où il se produit un relargage de phosphore, semble être confirmée par ces observations. Il est observé aussi une élimination de la DCO soluble dans la zone anaérobie qui est supérieure à celle théoriquement nécessaire pour la dénitrification des nitrates recyclés avec les boues décantées. Un exemple: le vingtième jour d'opération, la quantité de DCO soluble éliminée était

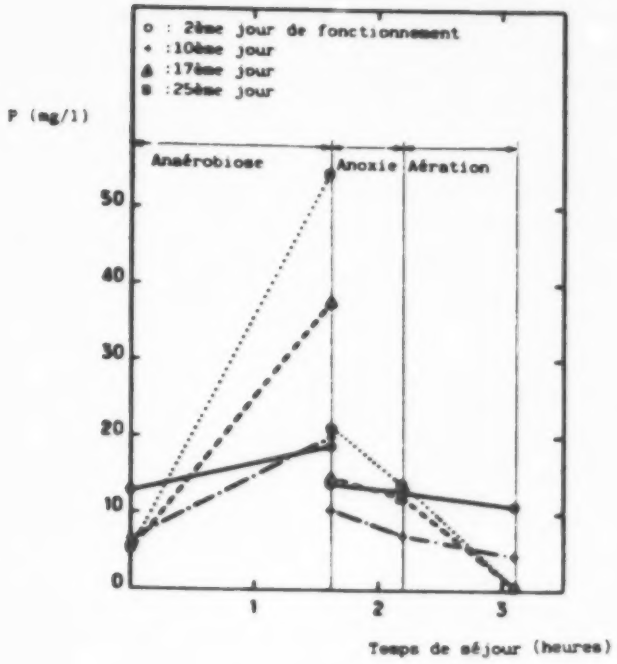

Fig. 8 : Evolution des concentrations en phosphore dans le pilote

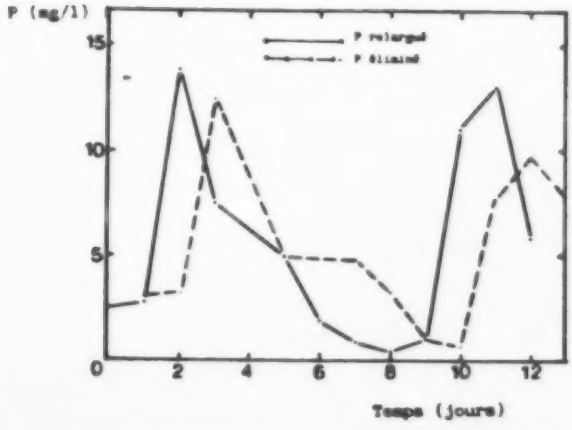

Fig. 9 : Relation entre le relargage du phosphore en anaérobiose et son élimination globale

222 - 97 = 115 mg/l. Pour la dénitrification, il fallait :
2 mg $N-NO_3$/l x 3,8 mg DCO/ mg $N-NO_3$ dénitrifié = 7,6 mg DCO/l. 107 mg DCO/l sont donc consommés indépendamment de la dénitrification.

Essais discontinus

Dans la description des réalisations existantes, il a été mentionné que la durée d'anaérobiose nécessaire dans le procédé "Phostrip" est sensiblement plus longue que dans les procédés du genre "Phoredox" et "UCT". Le fait que cette différence peut être attribuée à l'apport de matière organique, est illustré par les résultats d'un essai discontinu effectué sur la boue provenant de l'installation pilote (fig.10).

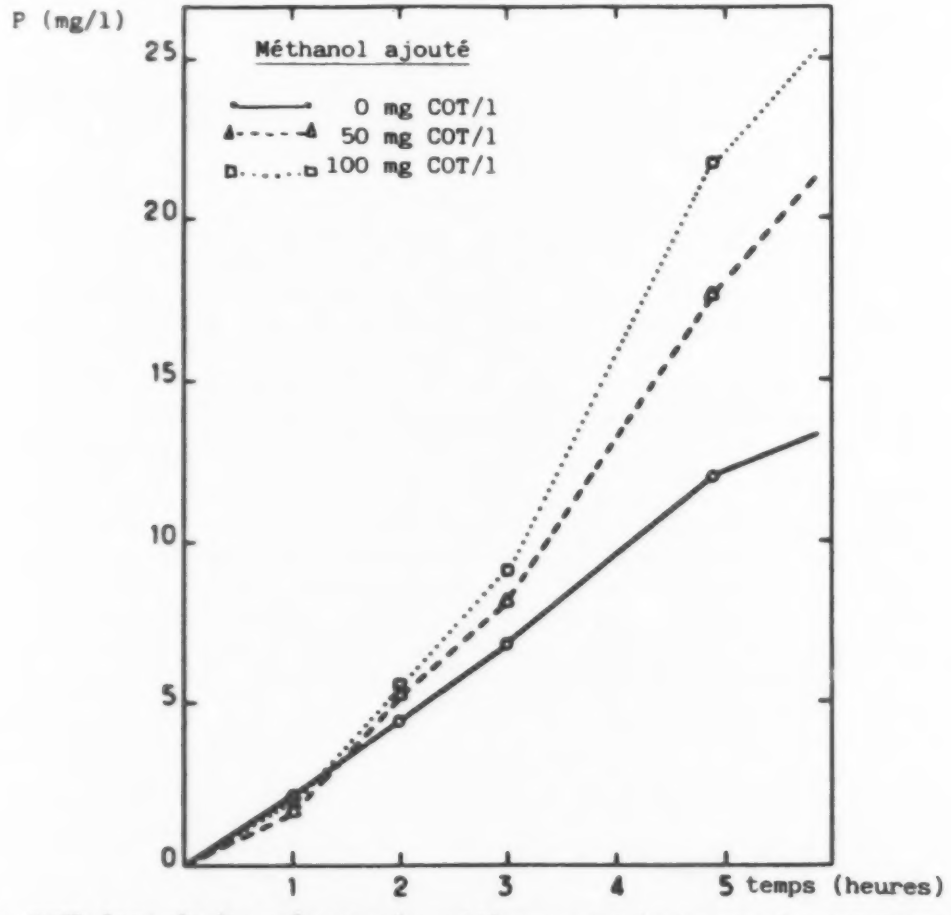

Fig. 10 : Influence de la présence de matières organiques sur le relargage du phosphore

Après aération pendant un temps équivalent au temps de séjour hydraulique dans le bassin aéré du pilote et après dénitrification, des quantités différentes de méthanol sont ajoutées aux échantillons de boue prélevés dans le bassins anoxique du pilote. On voit clairement que le relargage du phosphore est stimulé par l'ajout de matière organique. Cette influence devient plus importante avec des quantités ajoutées croissantes et se fait sentir surtout après des temps d'anaérobiose prolongés.

Si on utilise cependant de l'acétate comme source de carbone organique au lieu de méthanol, on constate que le relargage est immédiat et beaucoup plus rapide que dans le cas du méthanol (fig.11).

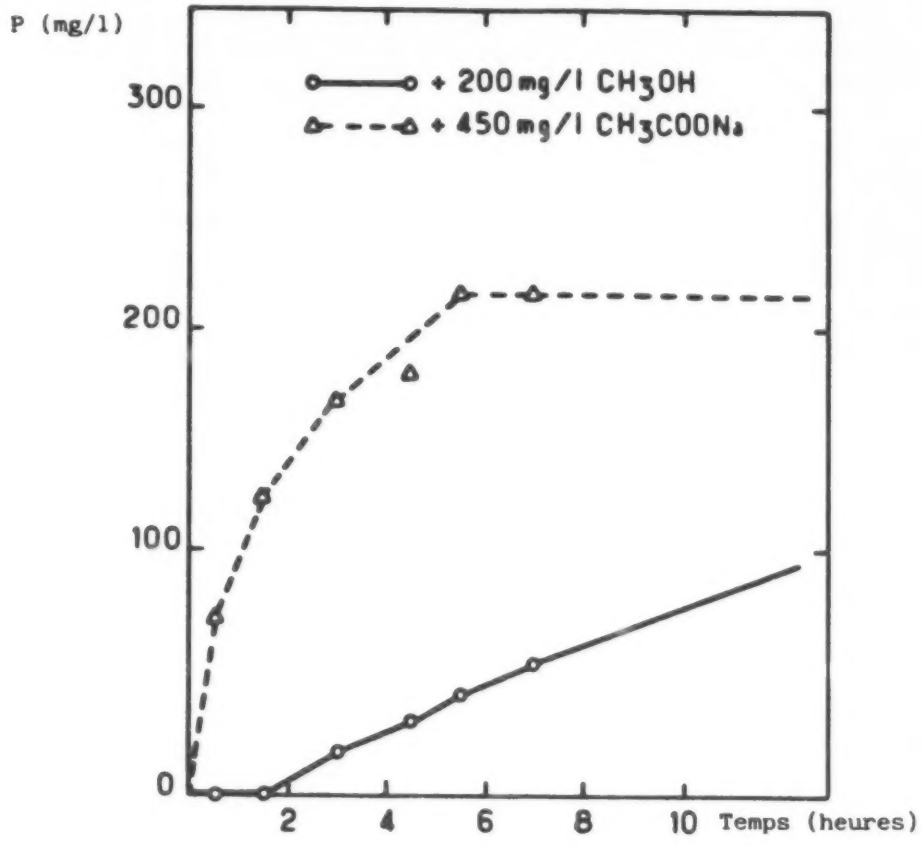

Fig. 11 : Influence de différentes matières organiques sur le relargage du phosphore

Après 30 minutes d'anaérobiose, la boue avait déjà relargué en solution 70 mg P/l, dans le cas où l'acétate avait été ajouté, tandis que le méthanol n'avait toujours pas induit un relargage mesurable. Après 6 h d'anaérobiose, la boue dopée en acétate s'était "vidée" du phosphore surstocké : la concentration en matières sèches était de 5,9 g/l, dont 6,1% de phosphore. Après avoir relargué 220 mg P/l, la teneur en phosphore dans la boue n'était plus que de 2,4 %, c'est-à-dire la valeur normale d'une boue activée classique.

Il est à noter que le relargage de phosphore sous l'influence d'acétate coincide avec une disparition de ce composé organique de la solution dans un rapport : 1 mole acétate consommée/1 mole P rélarguée.

ASPECTS MICROBIOLOGIQUES ET BIOCHIMIQUES

Ces observations concernant le rôle d'un substrat organique et de l'acétate en particulier, dans le relargage du phosphore, permettent de formuler une hypothèse quant aux mécanismes responsables de la déphosphatation

biologique.

Il est en effet possible que, dans des conditions anaérobies, des micro-organismes aérobies/anaérobies facultatifs dans la boue commencent à produire des acides gras volatils à partir de la matière organique présente dans l'eau (la vérification de cette possibilité est donnée par les résultats d'une incubation anaérobie d'une échantillon de boues sur un milieu nutritif après 2 jours à 37°C, il a été trouvé de l'acétate, du propionate et du butyrate en solution). L'essai discontinu précédent démontre que, en présence d'acides gras volatils (en particulier en présence d'acétate), un relargage de phosphore est induit dans une boue anaérobie provenant d'une installation "déphosphatante". Ce relargage consommant l'acétate au fur et à mesure de sa production, il est normal qu'il n'ait jamais été trouvé des acides gras en solution dans la zone anaérobie du pilote. La disparition d'une quantité de DCO soluble dans ce bassin pourrait cependant être expliquée par cette "acidogénèse".

L'acidogénèse n'est possible que dans un milieu à potentiel redox très faible. La présence de nitrates (accepteur d'électrons dans le métabolisme de certains micro-organismes) a tendance à augmenter le potentiel redox, donc à ralentir, voire bloquer la fermentation acide, ce qui inhibe le relargage du phosphore. L'apport de nitrates dans le bassin anaérobie d'une installation déphosphatante doit être limité à un minimum stricte, sinon les conditions d'anaérobiose ne seront plus assurées, et le rendement de déphosphatation diminuera. Les observations du rôle perturbateur des nitrates dans les stations "Bardenpho" et "Phoredox" peuvent facilement être expliquées par ce mécanisme. La figure 12 montre que, dans les essais discontinus, le relargage ne commence aussi qu'après dénitrification totale. Il a été observé cependant que l'ajout d'acétate peut induire un relargage même en présence de nitrates, ce qui tend à montrer que la présence d'acétate, plutôt que l'état d'anaérobiose en soi, est décisif pour le relargage du phosphore.

A : Période d'aération
B : Période d'anoxie
C : Période d'anaérobiose

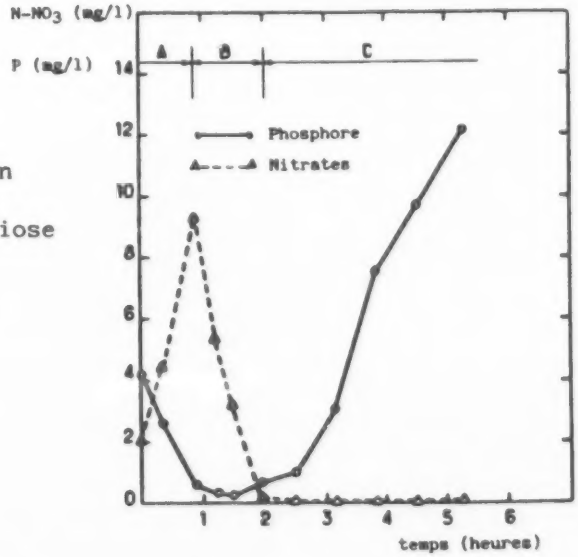

Fig. 12 : Influence des nitrates sur le relargage du phosphore

Pour expliquer le rôle particulier de l'acétate dans ces phénomènes, et la déphosphatation accrue dans des stations où le relargage en anaérobiose se produit, il faut entrer un peu plus en détail dans la microbiologie et la biochimie des organismes présents dans une boue activée. Il s'est avéré que certaines bactéries sont capables de former des granules de polyphosphates qui peuvent servir soit de réserve de phosphore pour des périodes de carence de cet élément, soit de reserves d'énergie : la liaison d'un groupe phosphate à la chaine de polyphosphates demande en effet une énergie de 7 kcal/mole, ce qui correspond à l'énergie qui est libérée lors de la dissociation d'ATP en ADP et phosphate.

Quand ces organismes sont placés en anaérobiose, il leur est possible de se servir de ces réserves d'énergie pour leur survie, et en plus, pour absorber du substrat organique. Dans le cas où l'acétate est présent (par exemple, par acidogénèse), un polymère organique riche en énergie peut être formé, aux dépens de l'énergie stockée dans les polyphosphates. Les réactions chimiques peuvent être écrites comme ci-dessous :

$$2 \text{ acétate} + 2 \text{ ATP} \longrightarrow \text{acétoacétate} + 2 \text{ ADP} + 2 \text{ Pi}$$

$$\text{acétoacétate} + \text{NAD}_{red} \longrightarrow \beta\text{ - hydroxybutyrate} + \text{NAD}_{ox}$$

Ce serait ce β -hydroxybutyrate qui est stocké sous une forme polymérisée, le PHB (poly-β-hydroxybutyrate).
On voit que pour chaque mole d'acétate transformé en acétoacétate, 1 mole d'ATP est consommée, donnant lieu au relargage de 1 mole de phosphates. Les réserves de polyphosphates seront hydrolysées afin de reconstituer l'ATP consommé.

Le PHB stocké sera ensuite métabolisé comme substrat organique dans les étapes aérobies et anoxiques du procédé. L'énergie libérée servira à la croissance bactérienne et à la reconstitution des polyphosphates. Le résultat global est une augmentation de la quantité de micro-organismes riches en phosphore (à cause des polyphosphates).

Dans une station déphosphatante, ce mécanisme permet d'expliquer, du moins qualitativement, la plupart des observations qui ont été faites : le relargage du phosphore et la consommation de matière organique en anaérobiose, l'influence des nitrates, la réabsorption finale du phosphore en aération. La possibilité de stocker du PHB pendant la phase anaérobie permet à ces micro-organismes d'épuiser déjà dans ce premier bassin une grande partie de la matière organique présente dans l'eau brute, ne laissant aux autres organismes que la partie la moins biodégradable de la pollution organique. On peut donc s'attendre à un enrichissement de ces organismes spécifiques, au détriment des organismes qui ne disposent pas de ce métabolisme particulier. Comme dans la phase d'aération, ces bactéries absorbent du phosphore pour faire des polyphosphates, la boue s'enrichit à ce moment-là en phosphore : des teneurs en P de l'ordre de 6 ou 7 % de la matière sèche sont couramment obtenues.

Plusieurs chercheurs ont essayé d'identifier les organismes responsables de ce mécanisme. Fuhs et Min Chen (1975) parlent de "souches ressemblant à des membres du groupe Acinetobacter - Moraxella - Mima".Deinema et al.(1980) sont plus affirmatifs : après isolement, toutes les souches représentant des granules de polyphosphates appartenaient au genre "Acinetobacter". Un isolement et une identification sur galeries API 20E des souches les plus nombreuses dans le pilote utilisé dans cette étude, a donné le résultat suivant :

Pseudomonas	: 21 %	des colonies apparues sur boîte de Pétri
Acinetobacter	: 18 %	" "
Aeromonas	: 14 %	" "

Il est à noter qu'Acinetobacter a l'habitude de se grouper en amas, de sorte que le nombre de colonies d'Acinetobacter apparues est certainement inférieur au nombre réel de ces bactéries dans l'échantillon de boues. Brodisch et Joyner (1982) n'ont retrouvé, cependant, que 5 % d'Acinetobacter dans deux boues déphosphatantes, parmi un nombre beaucoup plus élevé de Pseudomonas et d'Aeromonas. Elles en concluent qu'Acinetobacter ne peut pratiquement pas être la seule bactérie déphosphatante et que d'autres bactéries y participent.

Il a été prouvé qu'Acinetobacter est capable de stocker du phosphore en quantité supérieure aux besoins métaboliques. Des essais en culture pure, dans un milieu où l'acétate était la seule source de carbone, ont donné des teneurs en phosphore de l'ordre de 10 % du poids sec à la fin d'une culture où l'apport d'oxygène était limitant pour la croissance. Des essais dans des milieux sans acétate, ou avec apport d'oxygène non limitant, n'ont donné que des teneurs de 1 à 2,5 %, qui sont des valeurs normales pour une biomasse. Il paraît donc, qu'aussi bien la présence d'acétate que le manque d'oxygène soient nécessaires pour induire la surconsommation de phosphore par Acinetobacter. Il reste à examiner si, et dans quelles conditions, d'autres bactéries telles que Pseudomonas et Aeromonas, sont capables de s'enrichir en phosphore.

CONCLUSION

Des essais en laboratoire aussi bien que des réalisations industrielles ont démontré qu'il est possible d'éliminer, à des concentrations inférieures à 1 mg/l, le phosphore contenu dans une eau résiduaire urbaine, sans avoir recours à l'ajout de réactifs chimiques. Il est alors produit une boue en excès contenant 7 % de phosphore. Cet enrichissement en phosphore est réalisé par le développement de bactéries stockant des réserves de polyphosphates. La présence et la croissance exceptionnelle de ces bactéries (dont Acinetobacter est un représentant) est probablement favorisée par la transformation dans une phase anaérobie, de matière organique en acides gras volatils. Les organismes déphosphatants auraient la possibilité d'absorber et de stocker ces acides gras, même en anaérobiose, en utilisant pour celà l'énergie stockée dans les polyphosphates. La matière organique bien biodégradable de l'eau résiduaire ayant ainsi déjà été éliminée du milieu avant que les autres bactéries ne deviennent actives, les bactéries déphosphatantes se développent en plus grande quantité que dans une boue activée classique.

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COMPUTATION OF FRACTIONAL REMOVAL OF DISCRETE AND FLOCCULANT PARTICLES IN SETTLING BASINS

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ABSTRACT

In the conventional method of calculating the fractional removal of flocculant particles, it is assumed that the average settling velocity for different particle sizes in the lower depths in settling basins is the same as in the upper depths. This assumption is not quite correct, because the particles continuously flocculate and/or shear during the entire settling depth, and hence the average settling velocity changes both with depth and time.

In the modified method, a different approach for calculating the fractional removal of particles is proposed to overcome the deficiency in the conventional method. Instead of considering a vertical constant time line, a more reasonable horizontal constant depth line for that particular time is used. This approach is more reasonable because the settling velocities of different flocculating particles are all averaged over the same depth.

A typical example has been worked out to illustrate the applications of both the methods.

INTRODUCTION

In industrial and domestic wastewater treatment practices, settling tanks have been designed for the removal of both discrete and flocculant organic and inorganic suspended particles. Generally, the settling particles comprise of a broad spectrum of particles with different sizes and surface characteristics. In a near quiescent body of wastewater, larger particles settle at faster rates, thus overtaking finer particles in their descent. The flocculant particles may agglomerate, with a resultant increase in both the particle size and the settling velocity. Most settling in water and wastewater applications is of the flocculant type. In actual practice, the flocculation process is very complex due to hindered settling effects and also due to some shearing during settling of larger particles that had flocculated in the upper reaches of the settling tank. Thus, it is not always possible to predict the performance of sedimentation basins accurately for most water and wastewater applications from simple theoretical relationships.

The settling basins are designed to remove all the particles having a settling velocity equal to or greater than the design velocity. In addition, a certain fraction of smaller particles is also removed. In the conventional method (1,2) for calculating the fractional removal of flocculant particles, it is assumed that the average settling velocity for different particle sizes in the lower depths is the same as in the upper

depths, which is not quite correct. A different approach for calculating the fractional removal of particles is proposed to overcome the deficiency in analysis by the conventional method.

THEORETICAL BACKGROUND

Most settling in water and wastewater applications is of the flocculant type and follows either Newton's or Stoke's law; in the former case the settling velocity is related directly to the square root of the diameter of the particle, whereas in the latter case, directly to the diameter squared (1). In actual practice, the flocculation process is quite complex due to hindered settling effects and the particles may be moving in all directions prior to coalescence.

Evaluation of the sedimentation characteristics of a flocculant suspension can be accomplished by placing a quantity of water and wastewater in a column of approximately the same depth as the settling tank. The suspension is allowed to settle under quiescent conditions. The concentration of the particles is determined on samples withdrawn from several depths at different time intervals. The fraction of particles removed at each depth is then used to plot lines of equal fraction or percent removals, such as those illustrated in Fig. 1. Such lines are called 'Isoconcentration' (2) or Isopercentage (1) lines. These lines represent the maximum settling path for the indicated removal, and the depth to time ratio is equal to the minimum average settling velocity (2).

In Fig. 1, h_1, h_2, h_3, ... are the depths of the sampling ports from top; t_1, t_2, t_3, ... are the cumulative time between two consecutive samplings from the same port; R_{ma}, R_{mb}, R_{mc}, ... represent the fraction of suspended particles removed and these lines are called Isoconcentration and Isopercentage lines.

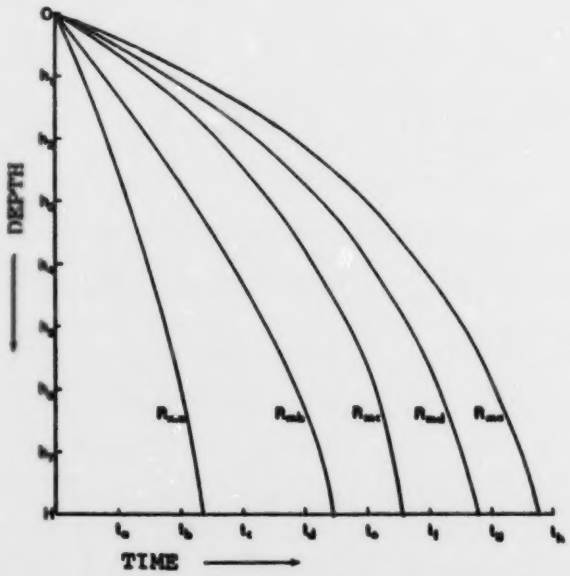

Fig. 1. Suspended Solids Removal vs Time and Depth

It can be seen from Fig. 1 that in a suspension when the particles flocculate during settling, their paths become curvilinear instead of a straight line assumed for discrete particles. The more pronounced the curvature of the settling paths, the greater the flocculation effect (3). It is emphasized that thermal currents will alter the settling velocities of the particles. In some cases slight temperature differences will result in apparently erratic settling curves. Domestic sewage appears to be particularly subject to this influence (3). Figure 1 clearly illustrates that the average settling velocity is varying both with time and depth.

ANALYSIS OF SUSPENDED SOLIDS REMOVAL FROM TIME-DEPTH DIAGRAM

The removal efficiency of flocculant particles in an ideal settling tank can be computed from the information provided in Fig. 2.

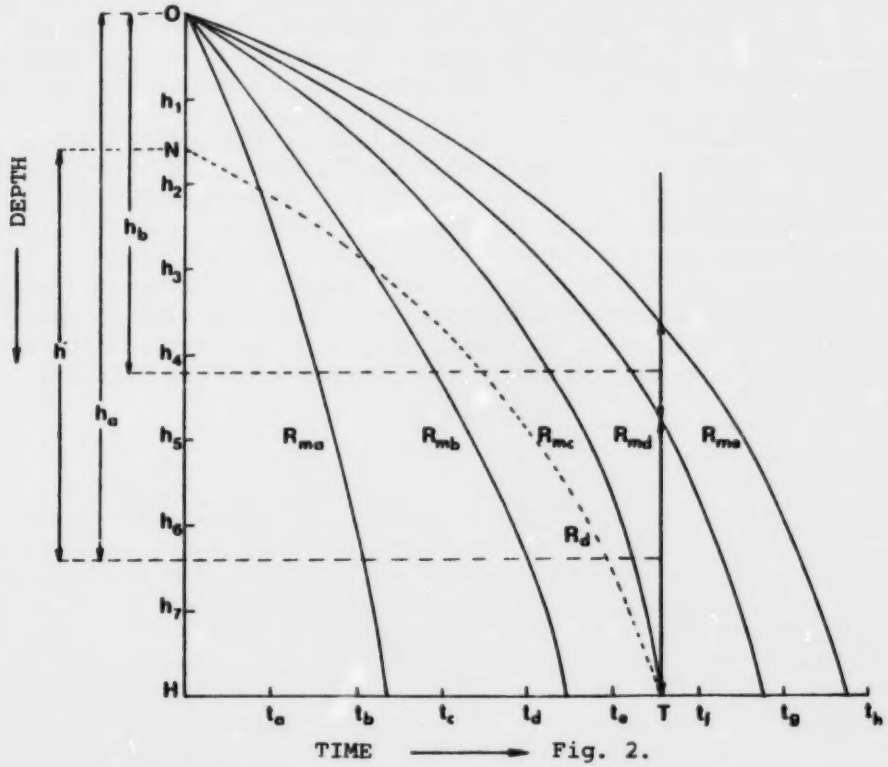

Fig. 2.

A. Conventional Method

Consider a tank that has an overflow rate of V_s and detention time T, where $V_s = \frac{H}{T}$, settling velocity averaged over depth H along settling path R_{mc}. All particles having an average settling velocity equal to or greater than V_s ($= \frac{H}{T}$) are assumed to be removed completely in a tank of depth H and in time T. Particles with smaller velocities are assumed to be removed in proportion to $\frac{V}{V_s}$, where V is less than V_s (2). From Fig. 2, it can be seen that a fraction of particles with settling paths of R_{ma} and R_{mb} will

be removed completely. Fraction R_{mc} also has settling velocity equal to V_s and will be removed in time T and depth H. The remaining fractions of solids between R_{mc} and R_{md}, R_{md} and R_{me}, and so on, have settling velocities less than V_s and will be removed in proportion to their settling velocities. Solids between isoconcentration lines R_{mc} and R_{md} are assumed to have an average settling velocity of $\frac{h_a}{T}$ and the solids between R_{md} and R_{me} have an average settling velocity of $\frac{h_b}{T}$. Their removal will be in the following proportions:

For solids between R_{mc} and R_{md}, fractional removal $= \frac{h_a}{T} / \frac{H}{T} = \frac{h_a}{H}$.

Similarly, for solids between R_{md} and R_{me}, fractional removal $= \frac{h_b}{T} / \frac{H}{T} = \frac{h_b}{H}$.

Consequently, an approximation for the overall removal R_m, is given by

$$R_m = R_{mc} + \frac{h_a}{H}(R_{md} - R_{mc}) + \frac{h_b}{H}(R_{me} - R_{md}) + \ldots.$$

The accuracy of analysis can be improved by adding more isopercentage lines and by decreasing the time interval (2).

In Fig. 2, the dotted curvilinear isopercentage removal line R_d is drawn parallel to removal line R_{md}. In this analysis, it is assumed that if R_{md} isopercentage line would have started from point N at height h' from bottom instead of point O at height H, 100% removal of those particles would have taken place. However, for a depth H and time T, the removal of particles having isopercentage removal line R_{md} will be fractional and in proportion to $\frac{h'}{H}$.

B. Modified Method

In the conventional method, while calculating fractional removal of particles, it is assumed that the paths followed by particles with isopercentage line R_{md} starting at O and N are parallel, Fig. 2. It is further assumed that solids between R_{mc} and R_{md} would settle at an average settling velocity of $\frac{h_a}{T}$. Similarly, solids between R_{md} and R_{me} would settle at an average velocity of $\frac{h_b}{T}$. Essentially, it is assumed that the settling velocity in the lower depths is the same as in the upper depths for different particle sizes. This is not quite true because the particles are continuously flocculating and/or shearing during the entire settling depth. Thus, the average settling velocity is not constant but varies with depth and time.

A different approach for calculating the removal of particles R_m is suggested in this paper to overcome the above mentioned deficiency in the analysis. For calculating the removal at some particular time T, it would be more precise to draw a line of total removal as shown in Fig. 3 by interpolation and then calculate the cumulative fractional removal of other particle sizes.

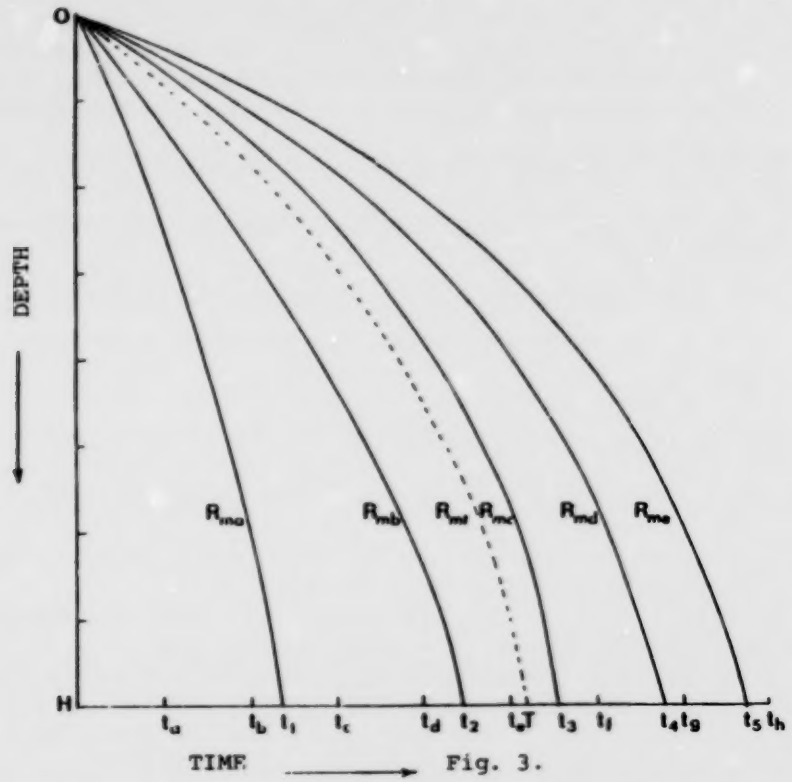

Fig. 3.

In this approach, instead of considering a vertical constant line for T, a more reasonable horizontal constant depth line for the given time T is used. Consequently, the total removal is given by:

$$R_m = R_{mt} + \left[\frac{H}{V_s} \Big/ \frac{1}{2} (T+t_3) \right](R_{mc} - R_{mt}) +$$

$$\left[\frac{H}{V_s} \Big/ \frac{1}{2} (t_3+t_4) \right](R_{md} - R_{mc}) +$$

$$\left[\frac{H}{V_s} \Big/ \frac{1}{2} (t_4+t_5) \right](R_{me} - R_{md}) + \ldots.$$

or $$R_m = R_{mt} + \frac{2T}{T+t_3} (R_{mc} - R_{mt}) + \frac{2T}{t_3+t_4} (R_{md} - R_{mc})$$

$$+ \frac{2T}{t_4+t_5} (R_{me} - R_{md}) + \ldots.$$

where T is the detention time, i.e. $T = \frac{H}{V_s}$.

This approach is more reasonable because the settling velocities of different flocculating particles are averaged over the corresponding depth.

Solids Removed at Indicated Depths, %

Time (min)	2 ft 0.610 m	3 ft 0.915 m	4 ft 1.220 m	5 ft 1.525 m	6 ft 1.830 m
0	0.0	0.0	0.0	0.0	0.0
5	17.0	14.0	10.0	9.0	8.0
10	28.0	23.5	19.5	17.5	15.0
20	41.5	36.0	30.5	27.0	26.5
30	54.0	48.0	41.0	37.0	33.0
40	62.0	54.5	46.5	44.0	41.5
50	66.5	60.0	54.5	50.5	46.0
60	73.0	66.0	58.5	55.5	52.5
75	75.0	71.0	66.7	62.5	58.0

TYPICAL EXAMPLE

A typical example has been worked out to illustrate the application of the proposed method. The data for this example have been taken from Table 3.3 in the 'Introduction to Wastewater Treatment Processes' by R.S. Ramalho (4) and are reproduced in Table I. Solids removed at depths of 3 ft (0.92 m) and 5 ft (1.53 m) are obtained by interpolation. The data along with isoconcentration lines are plotted in Fig. 4 and the calculations for total removal of particles by the two methods are shown below.

Conventional Method

For a detention period T = 35 min

$$\text{Total solids removal } R_m = 35 + \frac{1.700}{1.830}(40-35) + \frac{1.357}{1.830}(45-40) + \frac{0.991}{1.830}(50-45)$$

$$+ \frac{0.793}{1.830}(55-50) + \frac{0.633}{1.830}(60-55) + \frac{0.458}{1.830}(65-60)$$

$$+ \frac{0.351}{1.830}(70-65)$$

$$= 52.2\%$$

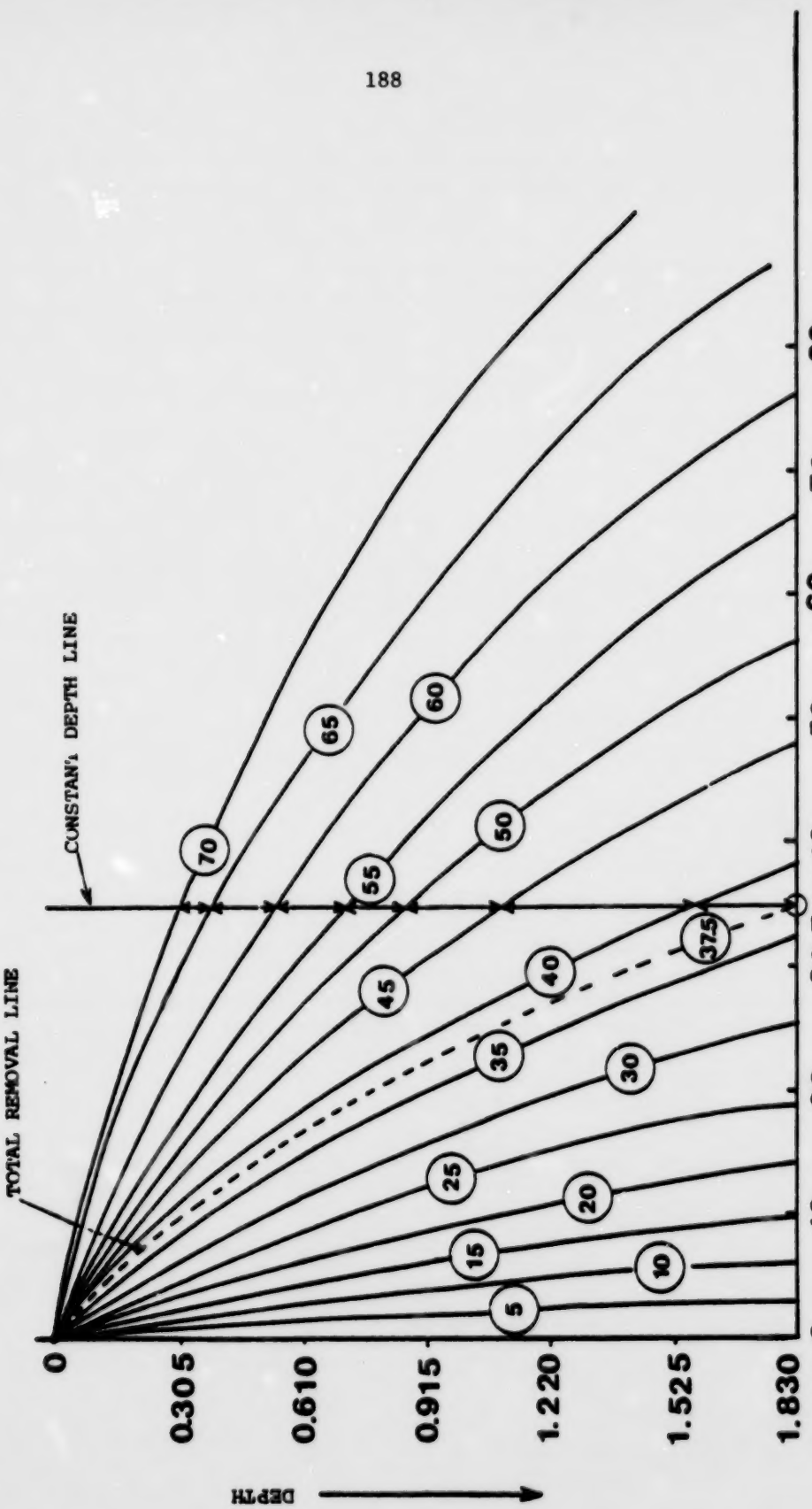
TOTAL REMOVAL LINE
CONSTANT DEPTH LINE
DEPTH
TIME
0
0.305
0.610
0.915
1.220
1.525
1.830
0
10
20
30
40
50
60
70
80
T=35
5
10
15
20
25
30
35
40
45
50
55
60
65
70
375

Fig. 4.

Modified Method

For T = 35 min

The total removal line for time T = 35 min is 37.5% removal line.

$$\text{Total solids removal, } R_m = 37.5 + \frac{2\times35}{35+38}(40-37.5) + \frac{2\times35}{38+48}(45-40)$$

$$+ \frac{2\times35}{48+56}(50-45) + \frac{2\times35}{56+66.5}(55-50) +$$

$$\frac{2\times35}{66.5+76}(60-55)$$

$$= 52.6\%$$

The subsequent terms for the fractional removal of particles have been ignored because they cannot be measured accurately from Fig. 4.

In the above example, the total removal of flocculant particles calculated by both the methods are different. In similar other examples also, the total removal of solids calculated by modified method were higher. As mentioned earlier, the modified method is more rational and as such the total removal of solid calculated by this method is expected to be more accurate.

CONCLUSION

The conventional method of calculating the total solids removal in a flocculant suspension is based on certain weak assumptions. Therefore, a more rational approach is proposed to overcome the deficiency in the conventional method. It is proposed to use a constant depth line for a particular time whereby the settling velocities of different flocculating particles are averaged over the same depth.

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MODÉLISATION DES PERFORMANCES DYNAMIQUES D'UNE USINE DE TRAITEMENT D'EAUX USÉES INDUSTRIELLES PAR BOUES ACTIVÉES

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RÉSUMÉ

L'étude entreprise concerne l'analyse des données journalières pour une période de trois années (de 1980 à 1982) pour une usine de traitement des eaux usées industrielles traitant des résidus phénoliques. Le but recherché a été de relier la réponse du procédé aux paramètres opérationnels par un modèle apte à traduire les variations temporelles de la biomasse et de la concentration en matières organiques du procédé.

Exprimant le substrat organique biodégradable par la variable DBO_5 et la biomasse par les SS un modèle différentiel traduisant un bilan de masse sur le système fut posé. Faisant intervenir la charge massique journalière, la recirculation et l'extraction des boues, il fut possible de calculer sur une base journalière le taux net d'assimilation de la matière organique R_s ainsi que le taux net de production de solides R_x par le système.

La démarche a consisté à comparer ces taux calculés à ceux obtenus de diverses formulations proposées dans la littérature afin d'en isoler le modèle le plus adéquat.

INTRODUCTION

La dynamique d'un système peut se définir comme la progression dans le temps des variables décrivant l'état de celui-ci en fonction de perturbations sur ses variables d'entrée ou ses paramètres de fonctionnement. La simple observation des valeurs décrivant l'état d'un procédé d'épuration biologique d'eaux usées, telle l'évolution de la concentration des micro-organismes ou de celle du substrat à la sortie indique la présence d'un phénomène transitoire. De plus, la variabilité encore plus prononcée des effluents industriels, tant dans leur volume que dans leur nature en fonction du temps, amplifie la dynamique des performances du système d'épuration.

Ainsi, dans le but de mieux contrôler le rendement de tels procédés, il devient essentiel d'apprécier et de caractériser les mécanismes de réaction mis en jeu. Une approche permettant de réaliser ceci consiste dans la pose d'un modèle mathématique faisant intervenir le plus adéquatement possible les variables opérationnelles du procédé [Busby and Andrews, 1975; Perdrieux and Thérien, 1980].

L'objectif de cette étude était de poser un tel modèle à partir de l'analyse de données provenant d'une usine traitant, par le procédé de boues activées, des eaux usées industrielles contenant principalement des résines, polymères, solvants et phénols.

Effectuant des bilans massiques différentiels sur les principales variables d'état du procédé, soit le substrat biodégradable et la biomasse, un modèle déterministe simple a été posé afin de décrire l'évolution temporelle de ces deux variables. Des simulations résultants de l'intégration de modèle sur ordinateur sont comparés aux profils observés expérimentalement et mettent en relief l'adéquation comparée de diverses expressions pour les taux d'assimilation du substrat et de la croissance de la biomasse. L'influence de la périodicité de certains facteurs environnementaux sur la réponse du modèle est également analysée.

DESCRIPTION DU PROCÉDÉ

Précédé d'un bassin d'égalisation de 2.46×10^6 ℓ (650,000 USG), le système d'épuration biologique considéré consiste de bassins combinés, l'un d'aération et l'autre de décantation. Le premier a un volume de 5.92×10^6 ℓ (1.56 MUSG) et le second de 0.13×10^6 ℓ (350,000 USG). Ces deux bassins sont couverts. Bien que protégés des intempéries, ils ne sont cependant pas chauffés l'hiver.

L'eau usée entre à une extrémité du bassin d'égalisation et est pompé de l'autre extrémité vers le bassin d'aération. Deux mélangeurs et deux diffuseurs d'air assurent l'agitation et l'apport en oxygène dissous au bassin d'aération. A la sortie, la liqueur mixte entre au bassin de décantation où les micro-organismes et la matière en suspension s'écoulent. Ceux-ci sont recirculés par entraînement au bassin d'aération par une ouverture au bas du décanteur.

L'eau clarifiée s'écoulant par trop-plein du décanteur est envoyée aux égoûts. Il est à noter qu'une certaine partie de cette eau clarifiée est recirculée occasionnellement au bassin d'égalisation et/ou au bassin d'aération lorsque le débit provenant de l'usine est peu élevé. Ceci a pour fonction principale de diminuer le temps de retention hydraulique des boues dans le bassin de décantation et ainsi d'empêcher un changement de nature de celles-ci, en boues plus difficilement décantables.

Les bassins d'aération et de décantation constituent les unités essentielles de ce procédé de boues activées et sont schématisés à la Figure 1. Le bassin d'égalisation joue un rôle important sur l'atténuation des fluctuations du débit et de la concentration de la liqueur à traiter. Cependant, aucune biodégradation significative ne se produit dans ce bassin et il n'est pas considéré davantage dans l'étude entreprise ici.

Il est à noter tout particulièrement, que l'extraction des boues du procédé se fait par pompage directement à partir du bassin d'aération et non dans la boucle de recyclage des boues comme pour la majorité des systèmes consistant de bassins d'aération et de décantation.

Les données disponibles pour ce procédé sont journalières et couvrent une période de trois années (1980-81-82). Les variables d'entrée au bassin d'aération sont le débit et la concentration de la liqueur à traiter exprimée par la DBO_5 et la DCO. La concentration des solides en suspension de la liqueur mixte dans ce bassin est également disponible. Les variables de sortie du bassin de

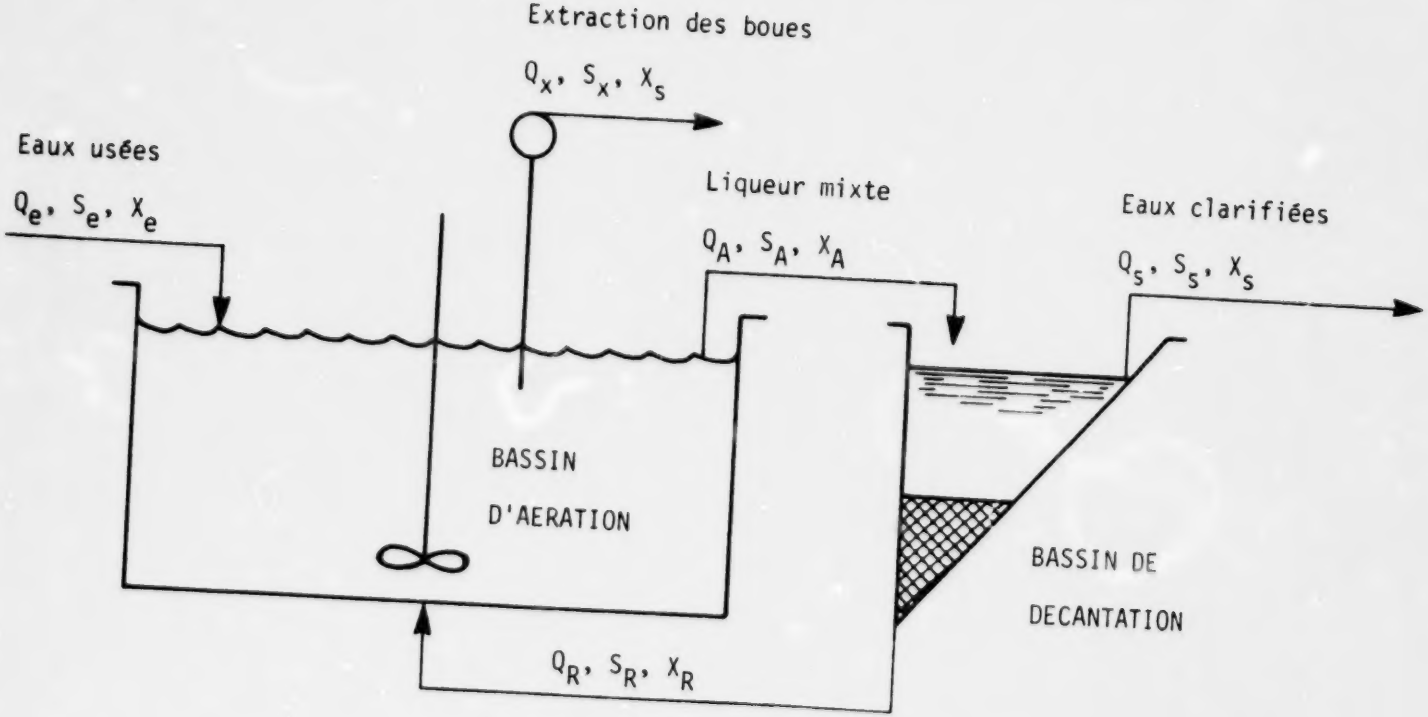


FIGURE 1: PROCEDE DE BOUES ACTIVEES

décantation sont la concentration des eaux clarifiées également exprimée par la DBO_5 et la DCO. La concentration en matières en suspension de cet effluent est également connue.

De plus, la concentration en oxygène dissous, le pH et la température en fonction du temps sont connus de façon journalière ainsi que le débit d'extraction des boues. Les Figures 2 à 5 montrent respectivement l'évolution de la moyenne hebdomadaire de la charge à l'entrée du bassin d'aération (Kg DBO_5/jour), de la concentration en matières biodégradables de l'eau clarifiée sortant du bassin de décantation (mg DBO_5/ℓ), la concentration des solides en suspension dans le bassin d'aération (mg/ℓ) et la température (°C) y existant.

Les fluctuations observées pour la charge, les solides en suspension et le substrat illustrent bien le caractère dynamique du système. Leur comportement ne semble pas présenter à priori de périodicité à l'exception de la température qui offre un cycle annuel ayant un maximum vers le 7ième mois (Figure 5). Les valeurs journalières moyennes calculées sur une période de 3 ans sont pour la charge de 1600 Kg DBO_5/jour et de 6550 mg/litre pour les solides en suspension dans le bassin d'aération.

Le temps de rétention moyen considérant le volume du bassin d'aération et le débit d'entrée moyen est de 12.6 jours.

D'autres informations concernant ce procédé sont disponibles ailleurs (Bellavance, 1983).

On peut également observer sur les Figures 3 et 4 que les variations temporelles pour les concentrations en substrat et en solides sont moins fluctuantes et présentent un caractère atténué par rapport à la charge à l'entrée (Figure 2).

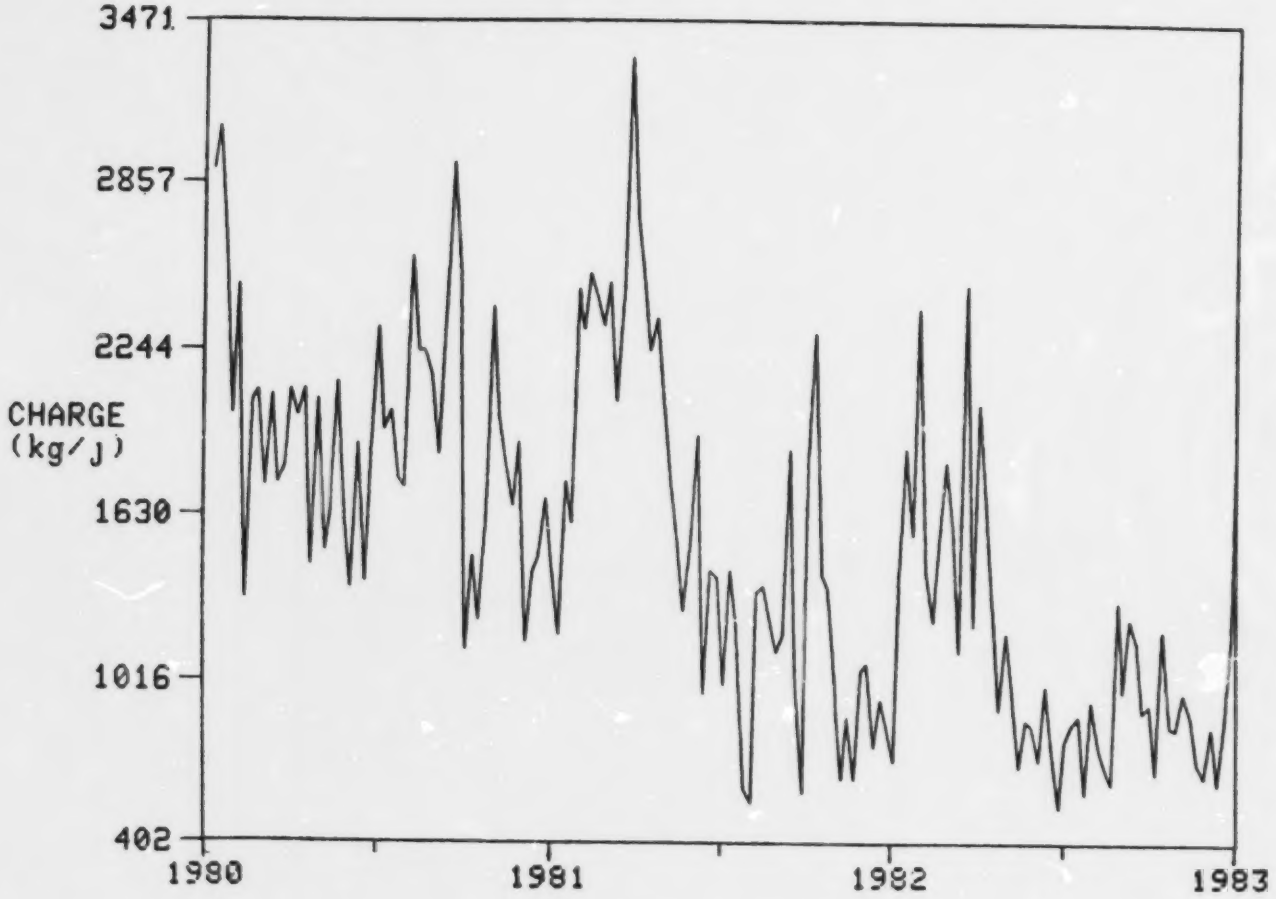
En comparant les Figures 2 et 3, on peut noter la correspondance existante entre la tendance à la hausse des solides en suspension au début des années 1981 et 1982, périodes pour lesquelles la charge est effectivement plus élevée. Ce phénomène est moins visible pour l'année 1980. Une comparaison semblable entre le profil de la charge (Figure 2) et celui de la concentration en substrat de l'effluent (Figure 4) est plus difficile à interpréter. Toutefois, il est possible d'observer pour chaque année qu'en réponse à une charge relativement accrue entrant au procédé pendant la période d'hiver l'effluent résultant est plus chargé en DBO_5. La comparaison des Figures 3 et 4 montre également une correspondance entre le rétablissement de la DBO_5 de l'effluent à une valeur plus basse au fur et à mesure que la concentration en solides, dans le système croit pendant ces périodes.

Ce que ces données illustrent finalement, c'est l'intéraction dynamique qui existe entre ces quantités dans le temps en plus de l'influence subie de facteurs environnementaux (température,...) et opérationnels (extraction des boues...).

POSE DU MODÈLE

La structure du modèle posé s'appuie sur un bilan massique différentiel effectué pour le substrat biodégradable et la biomasse du procédé de boues activées. Cette approche est relativement connue (Metcalf and Eddy, 1979; Benefield and Randall, 1980) et s'appuie sur un certain nombre d'hypothèses. Dépendant de l'hydraulique du procédé, de la nature de la liqueur à traiter (soluble, particulaire, mixte, ...) et de l'hétérogénéité de la culture consi-

CHARGE EN FCT(TEMPS)POUR 80-81-82 AVEC CHARGE EVALUEE / 1 SEMAINE.



1980-1981-1982(6 MOIS)

FIGURE 2

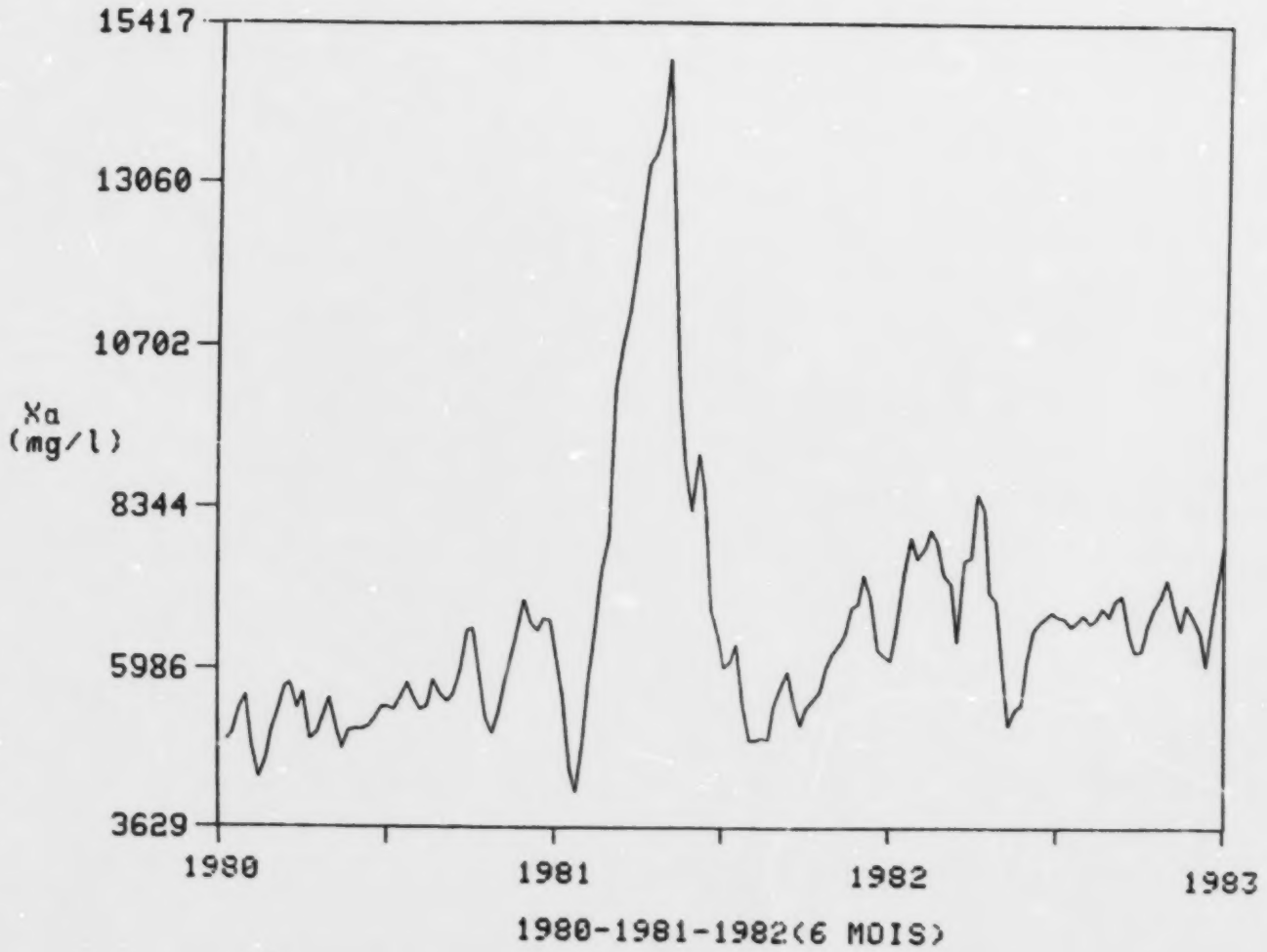

XA EN FCT(TEMPS)POUR 80-81-82 AVEC XA EVALUEE / 1 SEMAINE.
15417
13060
10702
8344
5986
3629
Xa
(mg/l)
1980
1981
1982
1983
1980-1981-1982(6 MOIS)

FIGURE 3

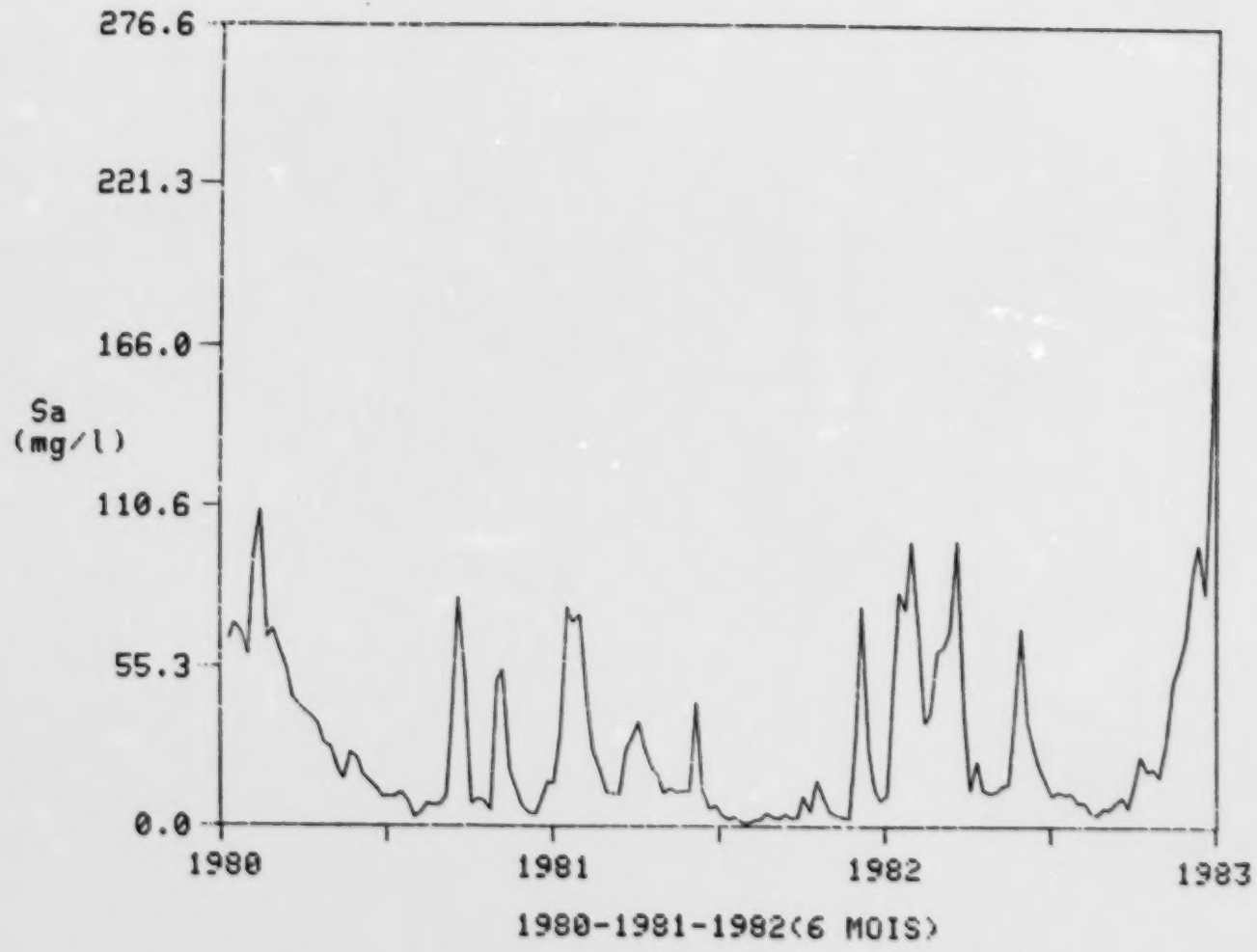
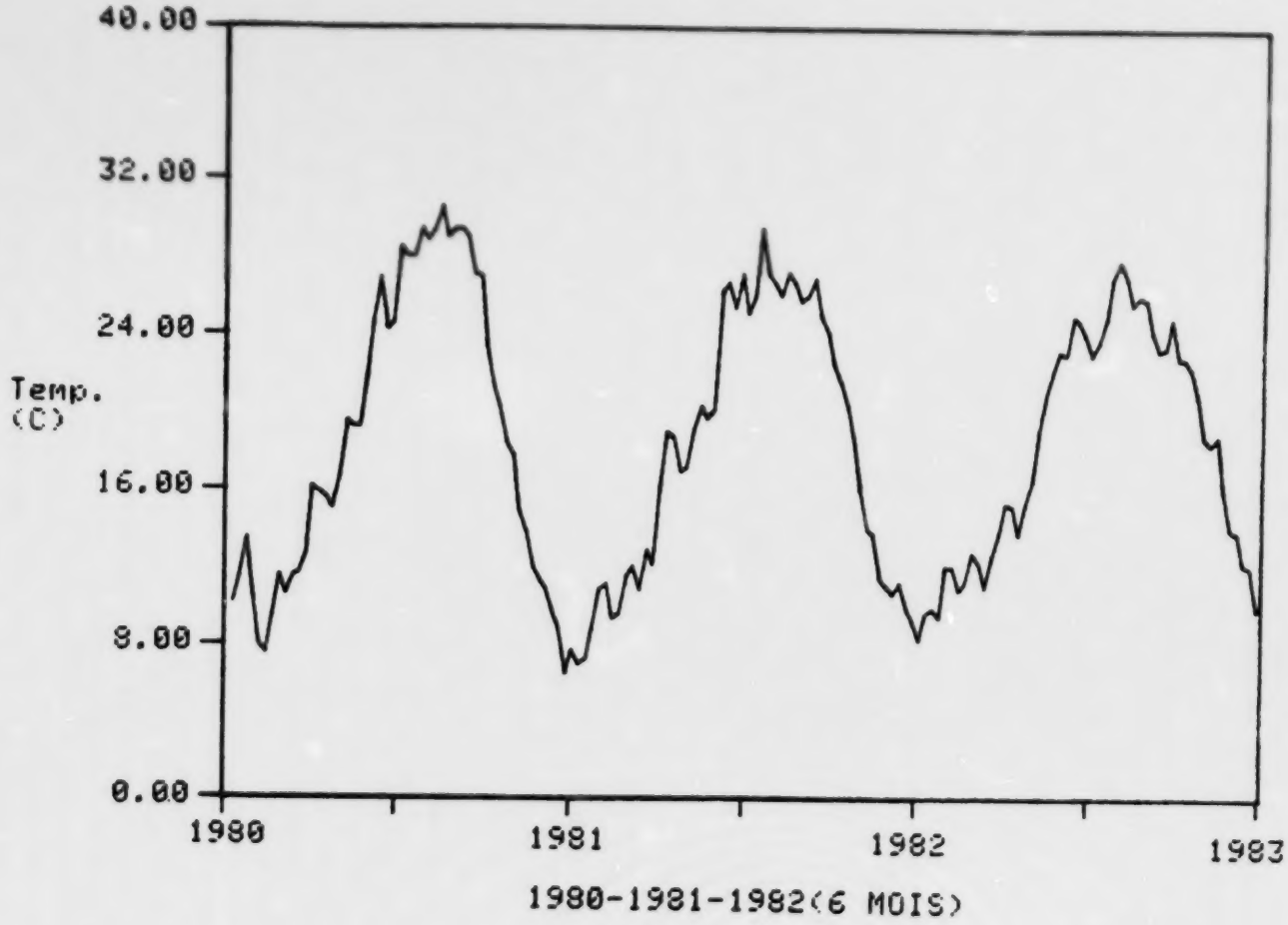

SA EN FCT(TEMPS)POUR 80-81-82 AVEC SA EVALUEE / 1 SEMAINE.
276.6
221.3
166.0
Sa
(mg/l)
110.6
55.3
0.0
1980
1981
1982
1983
1980-1981-1982(6 MOIS)

FIGURE 4

TEMPERATURE EN FCT(TEMPS)POUR 80-81-82 AVEC TEMP. EVALUEE / 1 SEMAINE.



1980-1981-1982(6 MOIS)

FIGURE 5

dérées, différents niveaux de complexité peuvent être obtenus pour un tel modèle (Ching, 1981; Cliff and Andrews, 1981).

L'hydraulique du procédé joue un rôle au niveau de la structure même du modèle mathématique alors que les diverses considérations touchant l'assimilation du substrat par les micro-organismes et leur croissance se traduisant par des formes plus ou moins compliquées des expressions pour les taux respectifs d'assimilation et de croissance.

Le bilan massique considère les quantités suivantes:

Accumulation = Entrée - Sortie ± Conversion

En accord avec le schéma du procédé de la Figure 1 et les termes y apparaissant pour le débit et les concentrations respectives pour le substrat et la biomasse, le bilan différentiel réalisé autour du bassin d'aération génère l'équation suivante:

$$V_A \frac{dS_A}{dt} = Q_e.S_e + Q_R.S_R - Q_x.S_x - Q_A.S_A - V_A.R_S \qquad (1)$$

et pour la biomasse:

$$V_A \frac{dX_A}{dt} = Q_e.X_e + Q_R.X_R - Q_x.X_x - Q_A.X_A - V_A.R_X \qquad (2)$$

Ici:

Q_e: débit volumétrique des eaux usées entrant au procédé (litre/jour)

Q_R: débit volumétrique de recirculation des boues activées de l'unité de décantation à l'unité d'aération (litre/jour)

Q_x: débit volumétrique d'extraction des boues à partir du bassin d'aération (litre/jour)

Q_A: débit volumétrique de la liqueur mixte sortant de l'unité d'aération (litre/jour)

V_A: volume du bassin d'aération (litre)

S_e: concentration du substrat biodégradable entrant au procédé (mg/litre)

S_R: concentration en substrat biodégradable des boues recirculées par entraînement entre l'unité de décantation et l'unité d'aération (mg/ℓ)

S_x: concentration en substrat biodégradable de la liqueur mixte extrait avec les boues (mg/ℓ)

S_A: concentration en substrat biodégradable entrant au procédé (mg/litre)

X_e: concentration de la biomasse dans les eaux usées entrant au procédé (mg/ℓ)

X_R: concentration de la biomasse dans les boues recirculées (mg/ℓ)

X_x: concentration de la biomasse extraite avec la liqueur mixte du bassin d'aération (mg/ℓ)

X_A: concentration de la biomasse dans le bassin d'aération (mg/ℓ)

R_S: taux d'assimilation du substrat par la biomasse dans le bassin d'aération (mg/ℓ-jour)

R_X: taux net de croissance de la biomasse dans le bassin d'aération (mg/ℓ-jour)

La pose des équations (1) et (2) font appels aux hypothèses suivantes:

(a) le bassin d'aération est parfaitement agité

(b) le volume du bassin d'aération est constant

Si l'on considère de plus les hypothèses suivantes:

(c) la concentration des micro-organismes présents dans la liqueur à traiter est négligeable devant celle existant dans le bassin d'aération, alors $X_e = 0$

(d) les caractéristiques de la liqueur extraites sont les mêmes que la liqueur mixte du bassin d'aération, alors: $S_x = S_A$ et $X_x = X_A$

(e) si la biodégradation du substrat n'a lieu que dans le bassin d'aération et que sa concentration n'est pas affectée par la décantation des solides, alors: $S_R = S_A$

Lorsque l'on considère également un bilan hydraulique en régime d'équilibre sur le bassin d'aération:

$$Q_e + Q_R = Q_x + Q_A \tag{3}$$

sur le procédé complet:

$$Q_e = Q_x + Q_S \tag{4}$$

et considérant que pour l'unité de décantation à l'équilibre:

$$Q_A.X_A = Q_R.X_R + Q_s.X_s \tag{5}$$

le modèle peut se réduire davantage sous la forme

$$V_A.\frac{dS_A}{dt} = Q_e.(S_e - S_A) - V_A.R_s \tag{6}$$

$$V_A.\frac{dX_A}{dt} = (Q_x - Q_e) \cdot X_s - Q_x \cdot X_A + V_A \cdot R_X \tag{7}$$

En général, les taux R_S et R_x sont des fonctions de Q_e et S_e dans le cas de procédé où le substrat est rapidement sorbé par la biomasse (Perdrieux, 1980) ou des fonctions de S_A et X_A sous d'autres considérations cinétiques (Harrington, 1976).

Lorsque les expressions pour R_s et R_x sont connues, l'intégration du modèle décrit par les deux équations différentielles (6) et (7) permet donc de simuler le comportement d'un système satisfaisant aux hypothèses émises plus haut. Les prédictions du modèle incorporent donc dans leur structure l'effet du débit Q_e de la charge massique ($Q_e.S_e$) ainsi que le taux d'extraction des boues Q_X. La connaissance de la variable X_S, la concentration de solides entraînés hors de l'unité de décantation exigerait, en théorie, la pose d'un modèle différentiel sur l'unité de décantation. Ces modèles sont en général complexes (équation différentielle partielle) et n'offrent pas des représentations généralement adéquates pour la dynamique de telles unités. Une approche alternative consiste à mesurer la sensitivité de la réponse de l'équation (7) à différentes valeurs que peut prendre X_S. C'est l'approche qui a été suivie ici (Bellavance, 1983). Elle a consisté dans un premier temps à utiliser une valeur constante pour X_S puis, dans un second temps, à utiliser le profil temporel de X_S mesuré à l'usine. Aucune différence significative n'ayant été observée dans les résultats obtenus pour X_A une valeur constante de $X_S = 65$ mg/ℓ pour les solides en suspension a été utilisée lors des simulations ultérieures.

Estimation des taux R_S et R_X

Utilisant un opérateur aux différences finies du premier ordre déphasé avant (Carnahan et al., 1969) pour représenter les opérateurs différentiels suivants:

$$\frac{dS_A}{dt} = \frac{S_A\,(t+\Delta t) - S_A(t)}{\Delta t}$$

$$\frac{dX_A}{dt} = \frac{X_A\,(t+\Delta t) - X_A(t)}{\Delta t}$$

les équations (6) et (7) deviennent des relations algébriques pour lesquelles les taux d'assimilation du substrat $R_S(t)$ et de croissance de la biomasse $R_X(t)$ peuvent être calculés à partir de données expérimentales.

Utilisant les données journalières (Δt = 1 jour) disponibles pour Q_e, Q_x, S_A, S_e, X_A et X_s, il fut possible de calculer les profils de $R_S(t)$ et de $R_X(t)$ en fonction du temps. Leur moyenne hebdomadaire est présentée aux Figures 6 et 7.

R_S représente le taux d'enlèvement du substrat ($mgDBO_5/\ell$) tandis que R_X représente le taux de variation net de la concentration en solides en suspension (mg/ℓ) dans le bassin d'aération. La grande similitude des profils temporels de R_S (Figure 6) et de la charge massique (Figure 2) indique une sorbtion quasi immédiate du substrat par la biomasse semblable à celle observée dans un procédé de contact-stabilisation (Metcalf and Eddy, 1979). Bien qu'un tel substrat sorbé sera éventuellement métabolisé et donnera lieu à une croissance de la biomasse, la comparaison de R_X (Figure 7) avec la charge massique (Figure 2) ne donne pas lieu à une relation aussi évidente. On note de plus des valeurs négatives de R_X à quelques occasions. Notant que ce taux réflète un taux net:

R_X = taux de croissance - taux de mortalité

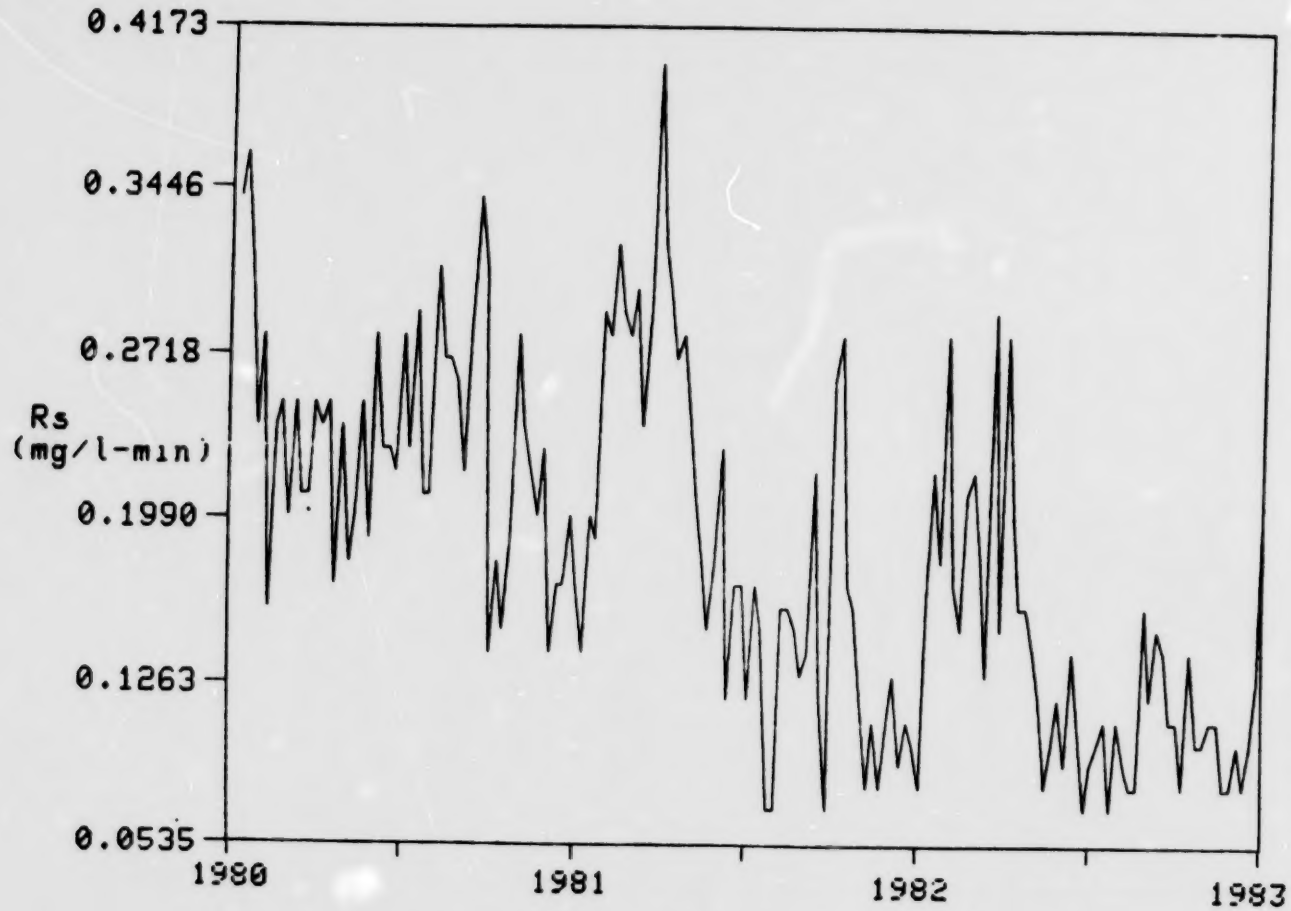
ceci pourrait s'expliquer par un phénomène de mortalité rapide d'une partie de la biomasse active provoquée, soit par la présence transitoire d'un élément toxique dans les eaux à traiter, soit par une limitation prolongée en substrat. Les résultats obtenus par calcul pour R_S et R_X permettent de comparer ces profils temporels avec ceux qui seraient obtenus par l'intermédiaire de formulations rapportées dans la littérature courante. La deuxième colonne de gauche du Tableau 1 rapporte un certain nombre de formulations plus ou moins courantes dans le but de comparer leur performance à représenter le plus adéquatement possible les profils expérimentaux des Figures 6 et 7.

Les formes A à D représentent toutes des expressions cinétiques limites de la forme E, soit le modèle de Monod (1949). Les formes F et G sont des expressions limites du modèle I compatible avec le concept de stockage et d'emmagasinement du substrat (Perdrieux et Thérien, 1980). La forme H est une forme quelque peu hybride.

Ici, les formulations du Tableau 1 pour R_X^* représentent en général des expressions pour le taux de croissance de la biomasse alors que R_X (Figure 7) représente un taux net. Toutefois, la considération d'un taux explicite de mortalité où:

$$R_{X^*net} = R_X^* - k_d \cdot X_A \qquad (10)$$

RS EN FCT(TEMPS)POUR 80-81-82 AVEC RS EVALUEE / 1 SEMAINE.



1980-1981-1982(6 MOIS)

FIGURE 6

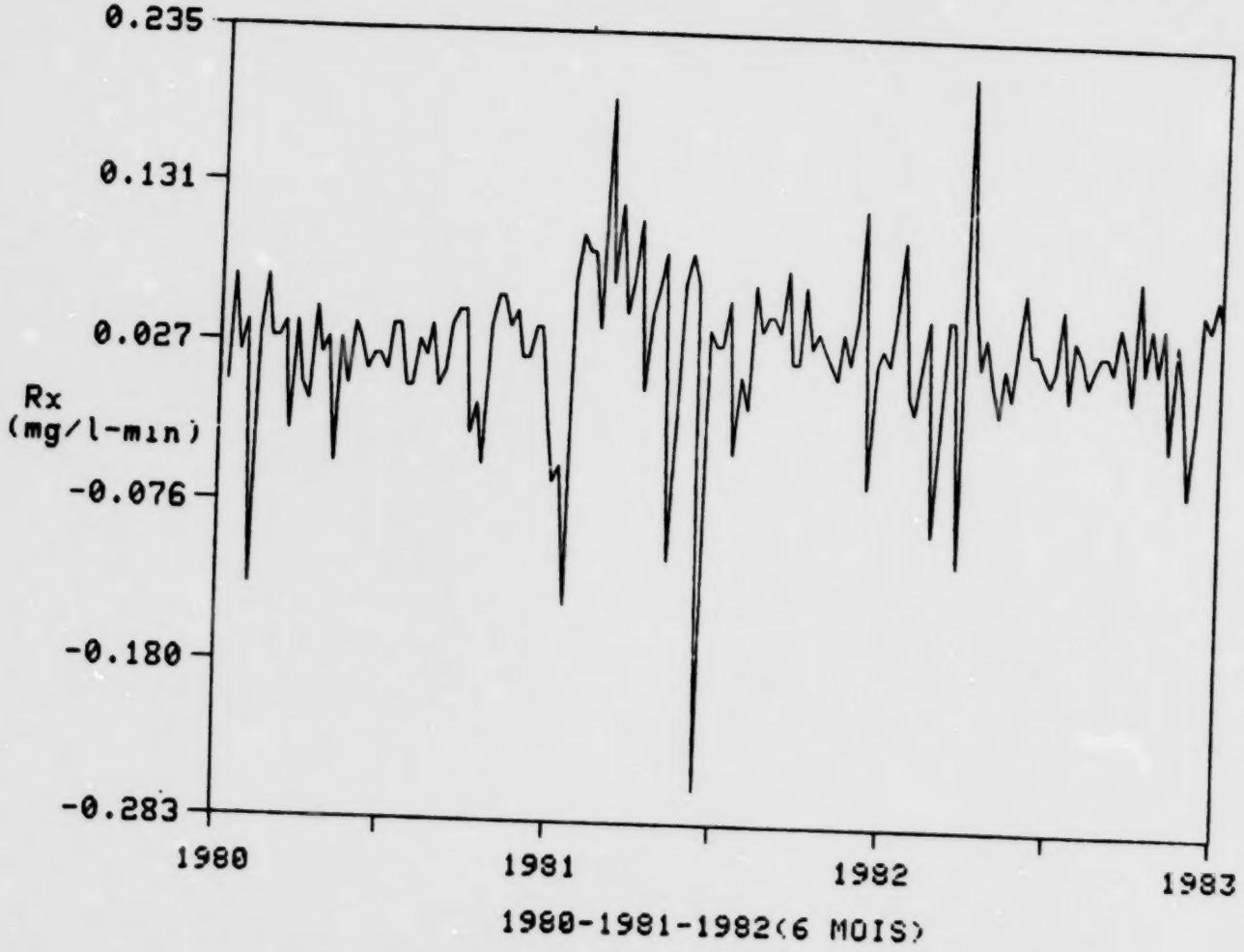

RX EN FCT(TEMPS)POUR 80-81-82 AVEC RX EVALUEE / 1 SEMAINE.
0.235
0.131
0.027
Rx
(mg/l-min)
-0.076
-0.180
-0.283
1980
1981
1982
1983
1980-1981-1982(6 MOIS)

FIGURE 7

s'est révélée en général non-significative à cause de la très faible valeur de ce dernier terme (Bellavance, 1983). En conséquence et désignant par R_S^* et R_X^* les formes du Tableau 1 (sauf la forme E) le calcul de K_i s'est effectué en faisant la moyenne arithmétique $\bar{K}_i$ des $K_i(t)$ journaliers calculés par:

$$K_i = R_i/(R_i^*/K_i) \qquad i = S,X \tag{11}$$

Les Figures 8 et 9 montrent respectivement les résultats obtenus lorsque R_S et R_X sont tracés pour une période de trois années en fonction de la charge massique (forme E du Tableau 1). Dans le premier cas, un coefficient de corrélation de 0.997 est obtenu et confirme l'observation faite auparavant entre les Figures 2 et 6.

A l'opposé une grande dispersion est observée dans le second cas avec une relation moins évidente sur une si grande période de temps.

Le calcul des $\bar{K}_i$ $i=S,X$ fut fait pour une période globale de 3 ans puis pour des périodes d'une année à la fois.

Simulation avec le modèle

Les valeurs moyennes ainsi obtenues pour des $\bar{K}_i$ $i=S,X$ furent utilisées avec les relations du Tableau 1 en calculant, à partir des données journalières pour Q_e, S_e, S_a et/ou X_A, les taux journaliers R_i^* $i=S,X$ dans chaque cas.

Le modèle décrit par les équations (6) et (7) fut donc utilisé pour simuler le comportement de S_A et X_A en fonction du temps pour chacune des formes du Tableau 1. Ces résultats furent par la suite comparés aux trajectoires expérimentales illustrées aux Figures 3 et 4.

L'adéquation des diverses formes pour R_i^* $i=S,X$ à bien décrire les tendances observées fut appréciée de façon visuelle tout en calculant, à chaque fois, le coefficient de variation α décrit par la relation suivante:

$$\alpha = \sum_{i=1}^{m} \left[\frac{(\text{valeur calculée} - \text{valeur observée})^2 * 100}{(n-1)\,(\text{moyenne des valeurs observées})} \right]^{\frac{1}{2}} \tag{12}$$

Cette mesure est adimensionnelle et est exprimée en % bien qu'elle puisse excéder 100%. Plus elle est petite plus le modèle est adéquat. Les résultats obtenus pour α sont présentés au Tableau 1 pour le calcul de X_A en colonnes 3 et 4 (utilisant R_x^*) et pour le calcul de S_A en colonnes 5 et 6 (utilisant R_s^*).

Ces résultats proviennent de l'intégration numérique des équations (6) et (7) par l'algorithme de Runge-Kutta du 4ième ordre (Carnahan et al., 1969). Divers pas d'intégration s'étalant de 0.002 à 1.0 jour furent utilisés afin de s'assurer de la justesse des résultats. Les formes pour R_s^* dont l'inadéquation se révéla par des valeurs extraordinairement grandes du coefficient de variation α (devenant ainsi une mesure peu significative) furent indiquées au Tableau 1 par un $\alpha > 1000$.

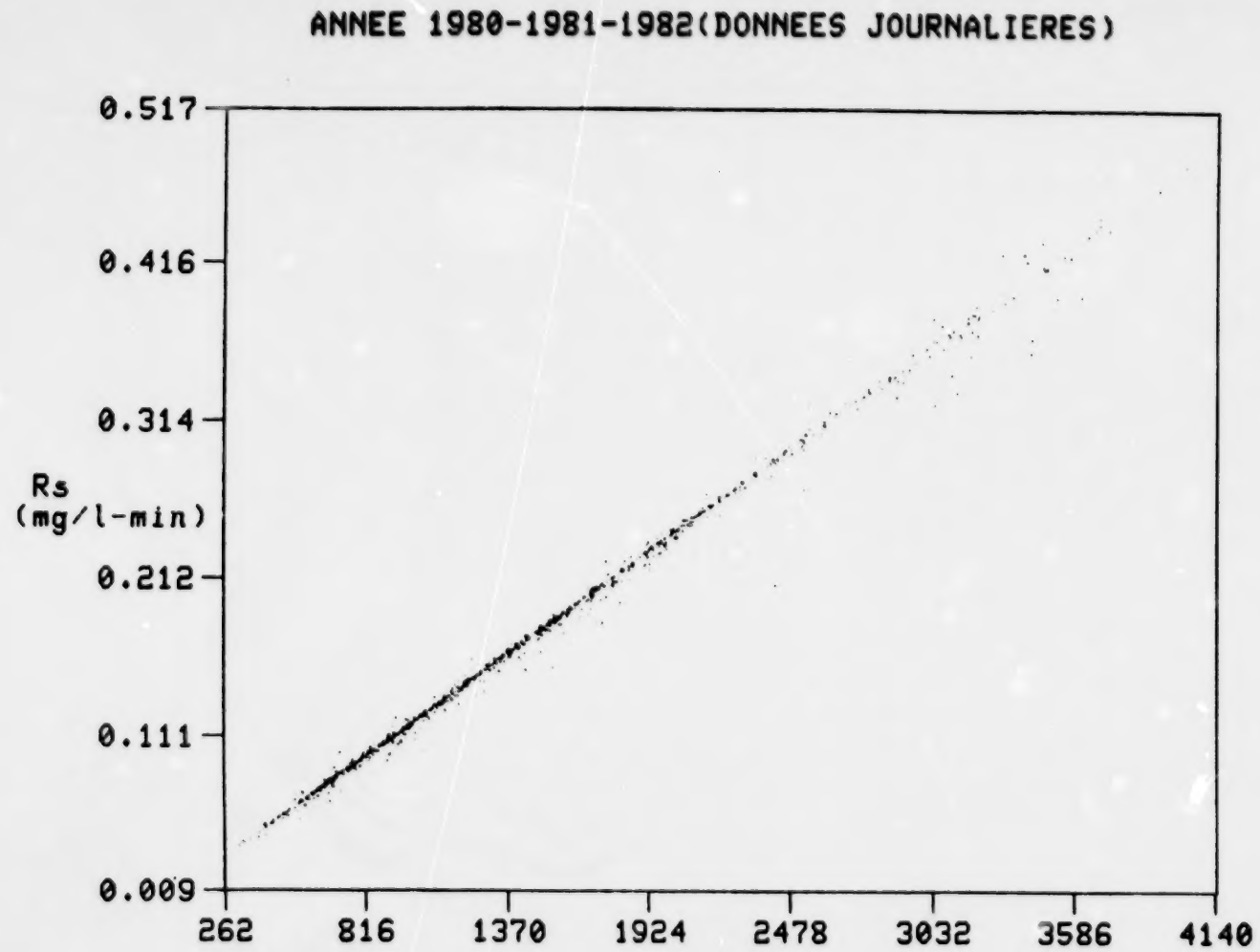

ANNEE 1980-1981-1982(DONNEES JOURNALIERES)
0.517
0.416
0.314
Rs
(mg/l-min)
0.212
0.111
0.009
262
816
1370
1924
2478
3032
3586
4140
CHARGE(SE1*QE1)(KG/J)

FIGURE 8

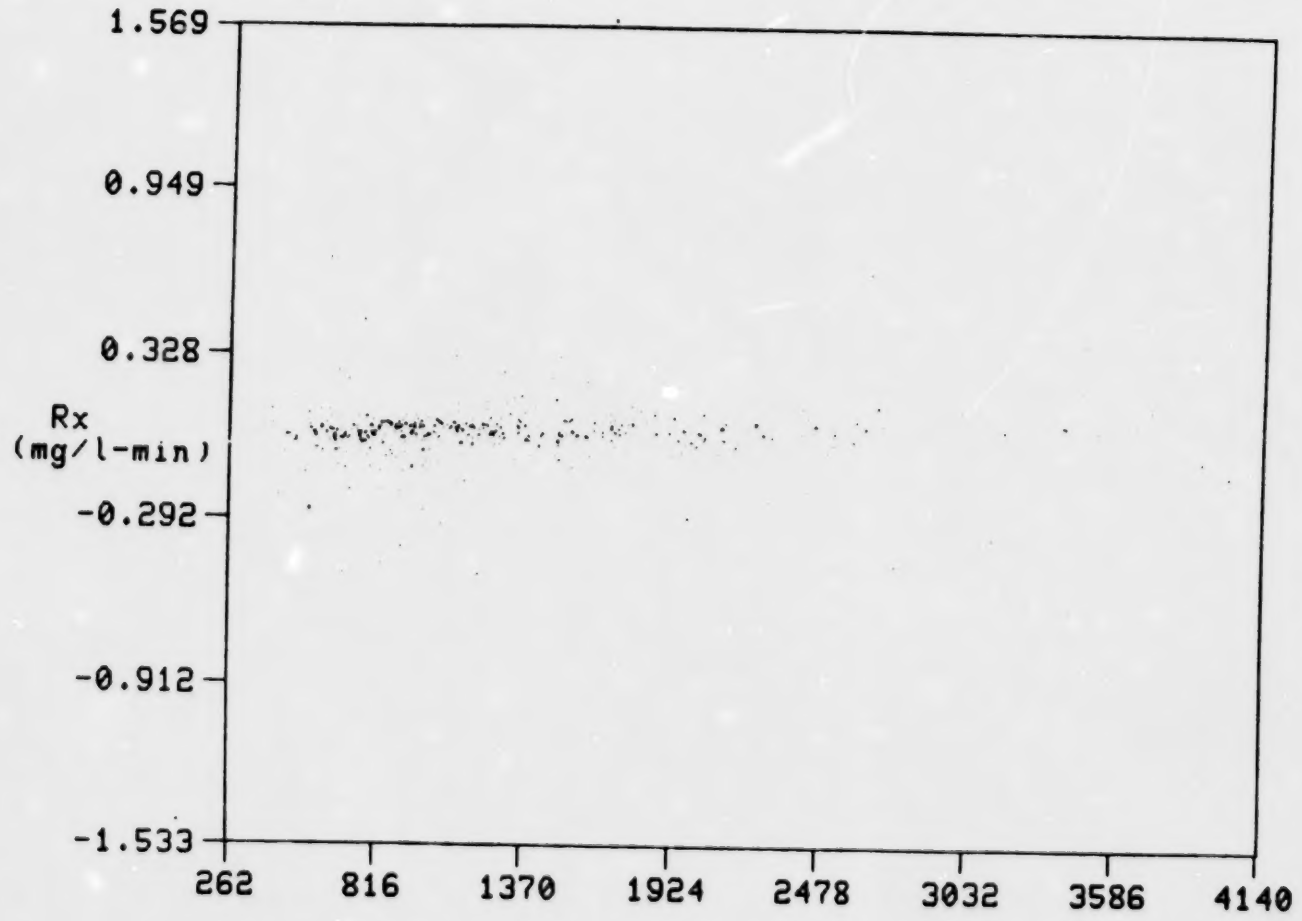

ANNEE 1980-1981-1982(DONNEES JOURNALIERES)
1.569
0.949
0.328
Rx
(mg/l-min)
-0.292
-0.912
-1.533
262
816
1370
1924
2478
3032
3586
4140
CHARGE(SE1*QE1)(KG/J)

FIGURE 9

TABLEAU 1: RÉSULTATS DE SIMULATION POUR DIFFÉRENTES FORMULATIONS DE R_s* et R_x*

		X_A		S_A	
Forme	R_i*, i = s,x	Coefficient de variation (%) (K_x sur 3 ans)	(K_x annuel)	Coefficient de variation (%) (K_s sur 3 ans)	(K_s annuel)
A	K_i	26.7	15.8	> 1000	> 1000
B	$K_i \cdot S_A$	40.5	31.2	143	140
C	$K_i \cdot X_A$	28.6	16.2	> 1000	> 1000
D	$K_i \cdot X_A \cdot S_A$	43.0	34.8	172	165
E	$\frac{K_i \cdot S_A \cdot X_A}{K_n + S_A}$	30.4	18.9	> 1000	> 1000
F	$K_i \cdot S_e$	30.9	16.1	> 1000	> 1000
G	$K_i \cdot Q_e$	30.1	16.1	> 1000	> 1000
H	$K_i \cdot S_e \cdot X_A$	32.9	23.5	> 1000	> 1000
I	$K_i \cdot S_e \cdot Q_e$	34.2	16.7	134	127

DISCUSSION DES RÉSULTATS

Les résultats obtenus pour la formulation de R_x indiquent que plusieurs expressions cinétiques produisent des résultats aussi valables les uns que les autres en regard du coefficient de variation. C'est le cas des formes A, C, E, F, G et I. Par contre, les formes B et D sont significative moins adéquates. Dans le cas de formulations pour R_s une discrémination plus évidente est observée avec les formes B, D et I qui offrent des performances très supérieures aux autres. La forme I avec $R_s = K_s \cdot Q_e \cdot S$ produit les meilleurs résultats.

Ce résultat, avec une forme simplifiée du modèle de stockage (Perdrieux, 1977), n'est pas étranger au fait que dans le phénomène d'emmagasinement du substrat par les cellules le taux d'assimilation R_s est proportionnel, dans une large mesure, au taux d'apport en matières organiques biodégradables au système (représenté par la charge $Q_e \cdot S_e$).

Dans un procédé de boues activées où en général l'apport en substrat constitue le facteur limitatif à la croissance de la biomasse, le taux R_x de cette croissance risque également de dépendre du taux auquel le substrat est acheminé à l'intérieur des cellules. Conséquemment, le taux R_x serait également proportionnel à R_s ou à la charge organique $Q_e \cdot S_e$ sous des conditions limitantes en substrat. Dans les moments où des apports substantiels en substrat au système donnent lieu à l'emmagasinement continu de substrat dans les cellules, le taux de croissance devient relativement indépendant du taux de stockage ou même de la quantité stockée au delà d'un seuil minimal. C'est ce

qui pourrait expliquer pourquoi, sur des périodes d'une année ou de trois années, où une alternance de ces conditions surviennent, les résultats du Tableau 1 montrent des résultats aussi satisfaisants d'une forme $R_x = \bar{K}_x$ ou $R_x = \bar{K}_x.Q_e.S_e$. La première forme pourrait approximer le comportement du phénomène de croissance sous des conditions it d'abondance en substrat, le taux R_x n'étant pas proportionnel à la quantité de substrat stockée mais représentant le taux maximal de croissance K_x sous ces conditions. La deuxième forme serait plus réaliste sous des conditions limitatives en substrat.

La Figure 10 illustre le résultat de la simulation avec le profil temporel observé à la station d'épuration sur une période de trois années lorsque $R_x = \bar{K}_x.Q_e.S_e$ où $\bar{K}_x$ est ajusté annuellement.

Pour 1980, 1981 et le premier trimestre de 1982, le modèle simule raisonnablement bien le comportement dynamique du système avec l'apparition des pics significatifs aux bons moments avec une ampleur bien rendue. Toutefois, les résultats sont moins satisfaisants pour les deux autres trimestres de l'année 1982.

La Figure 11 présente les résultats de la simulation avec $R_x = \bar{K}_x$ où $\bar{K}$ est ajusté annuellement. Bien que dans ce cas l'année 1982 soit bien représentée, ce modèle ne peut reproduire adéquatement le pic observé au printemps de l'année 1981.

La Figure 12 représente le comportement simulé du modèle utilisant $R_s = \bar{K}_s.Q_e.S_e$ avec un paramètre $\bar{K}_s$ ajusté annuellement. Les résultat indiquent en général un manque à reproduire les pics ponctuels de l'année 1980 bien que l'ampleur de ceux du début de l'année 1982 soit simulée. Un décalage relativement important est noté vers l'été et vers la fin de l'année 1982. Le Tableau II résume la valeur des paramètres $\bar{K}_x$ et $\bar{K}_s$ utilisé de ces simulations.

Les résultats sont jugés satisfaisants compte-tenu qu'ils utilisent des paramètres $\bar{K}_s$ et $\bar{K}_x$ constants sur des périodes consécutives d'une année. En effet, les paramètres liés à l'assimilation du substrat et à la croissance de la biomasse dans un système réel dépendent en général de la nature de l'affluent à traiter ainsi que des conditions environnementales et sont, de ce fait, variables en fonction du temps.

Connaissant, par calcul, les valeurs journalières des taux R_s, R_x ainsi que des paramètres K_s et K_x une analyse statistique de leur fluctuations temporelles fut faite afin de déceler si des cycles particuliers pouvaient s'en dégager. A cet effet, le calcul de la densité spectrale de R_s, R_x, K_x et K_s fut faite. Le tracé de celle-ci en fonction du temps permet d'observer ou non l'existence et l'importance de périodes caractéristiques distinctes d'une série temporelle de valeurs discrètes (Malivaud, 1978; Delong et al., 1980). Les résultats obtenus montrèrent que R_s, R_x, K_x ainsi que
et al., 1980). Les résultats obtenus montrèrent que R_s, R_x, K_x ainsi que la charge massique $Q_e.S_e$ entrant au procédé n'ont aucune période significative

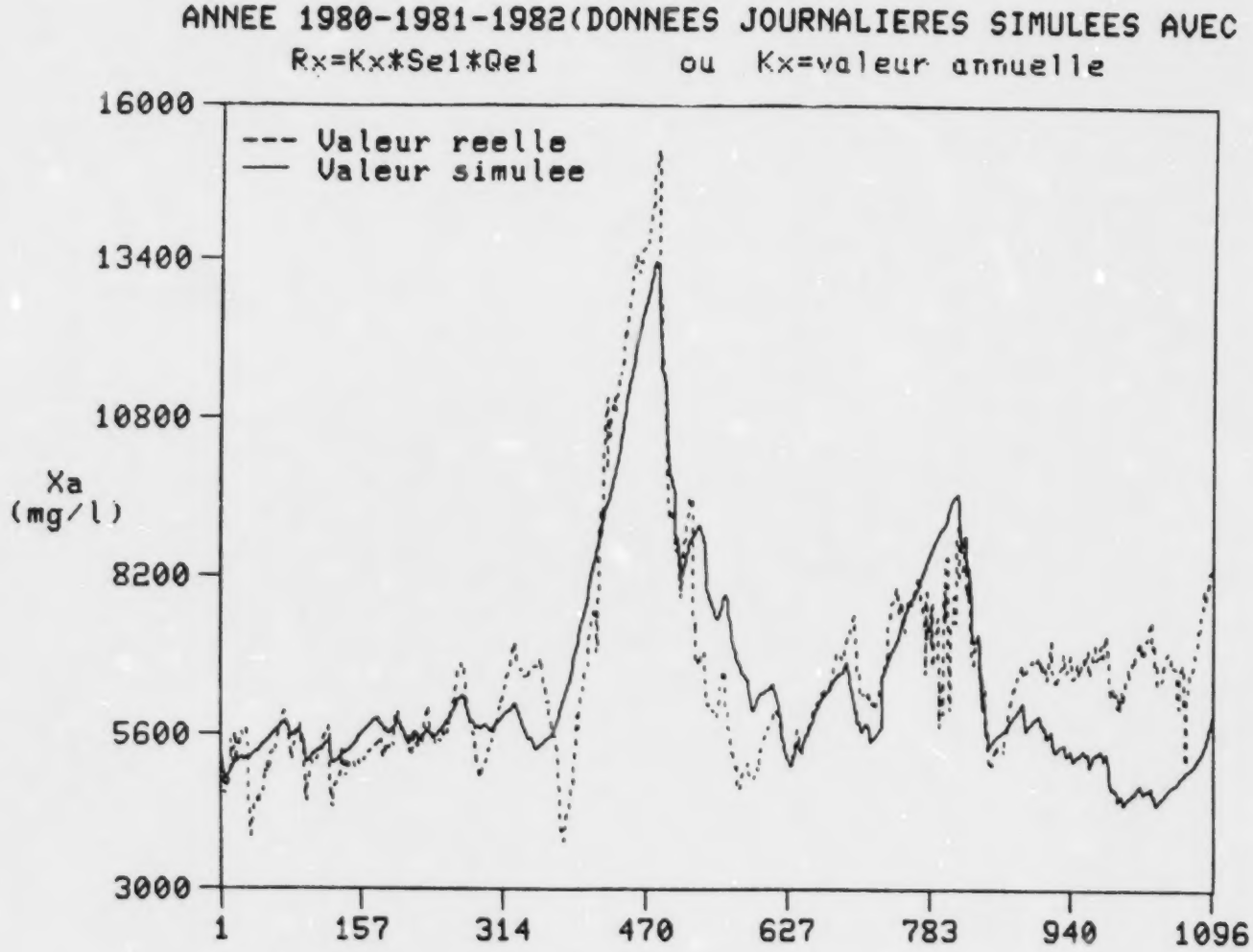

ANNEE 1980-1981-1982(DONNEES JOURNALIERES SIMULEES AVEC
Rx=Kx*Se1*Qe1 ou Kx=valeur annuelle
--- Valeur reelle
—— Valeur simulee
16000
13400
10800
8200
5600
3000
Xa
(mg/l)
1
157
314
470
627
783
940
1096
1980-1981-1982(JOUR)

FIGURE 10

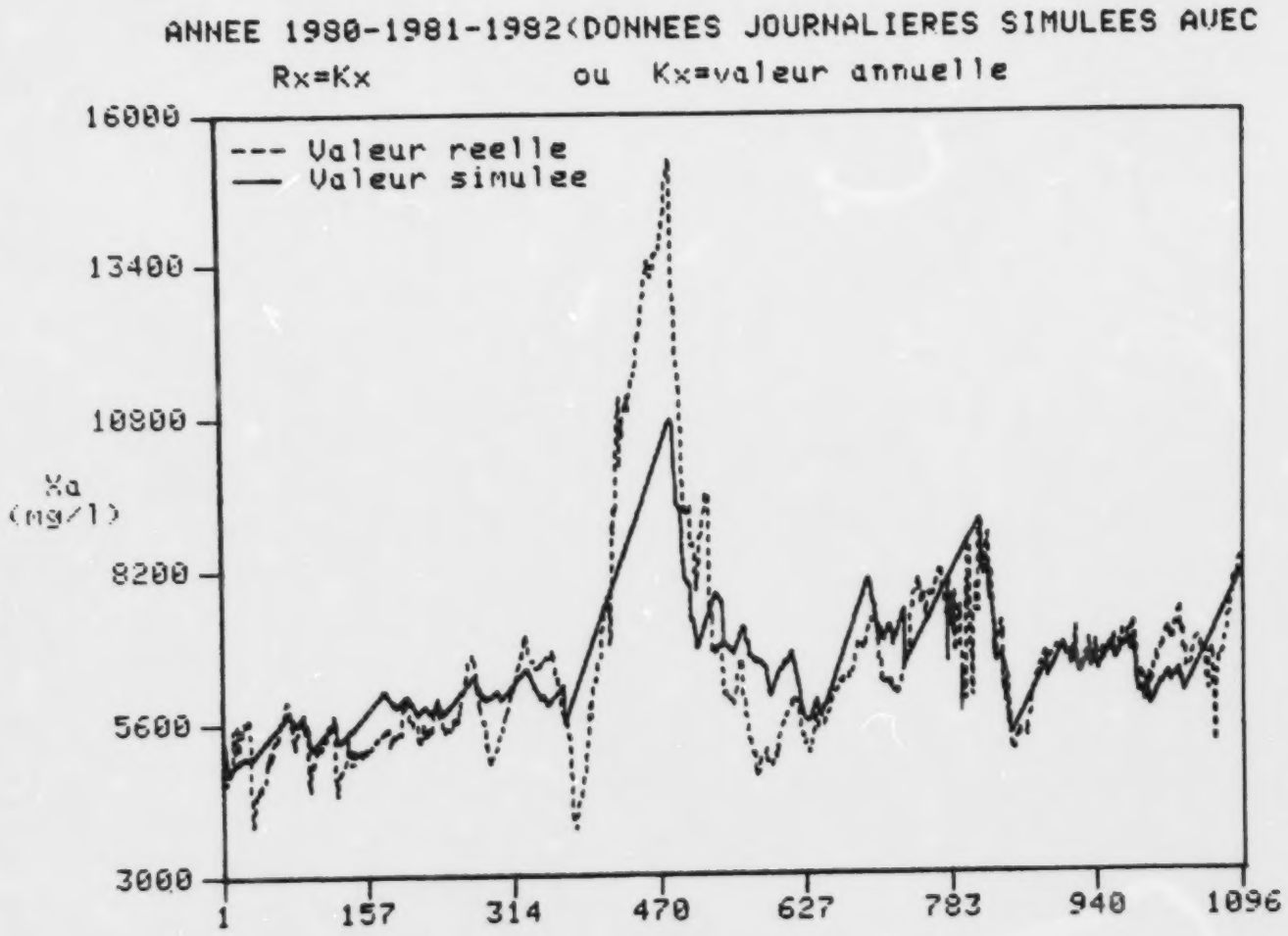

ANNEE 1980-1981-1982(DONNEES JOURNALIERES SIMULEES AVEC
Rx=Kx ou Kx=valeur annuelle
--- Valeur reelle
—— Valeur simulee
16000
13400
10800
8200
5600
3000
Xa
(mg/l)
1
157
314
470
627
783
940
1096
1980-1981-1982(JOUR)

FIGURE 11

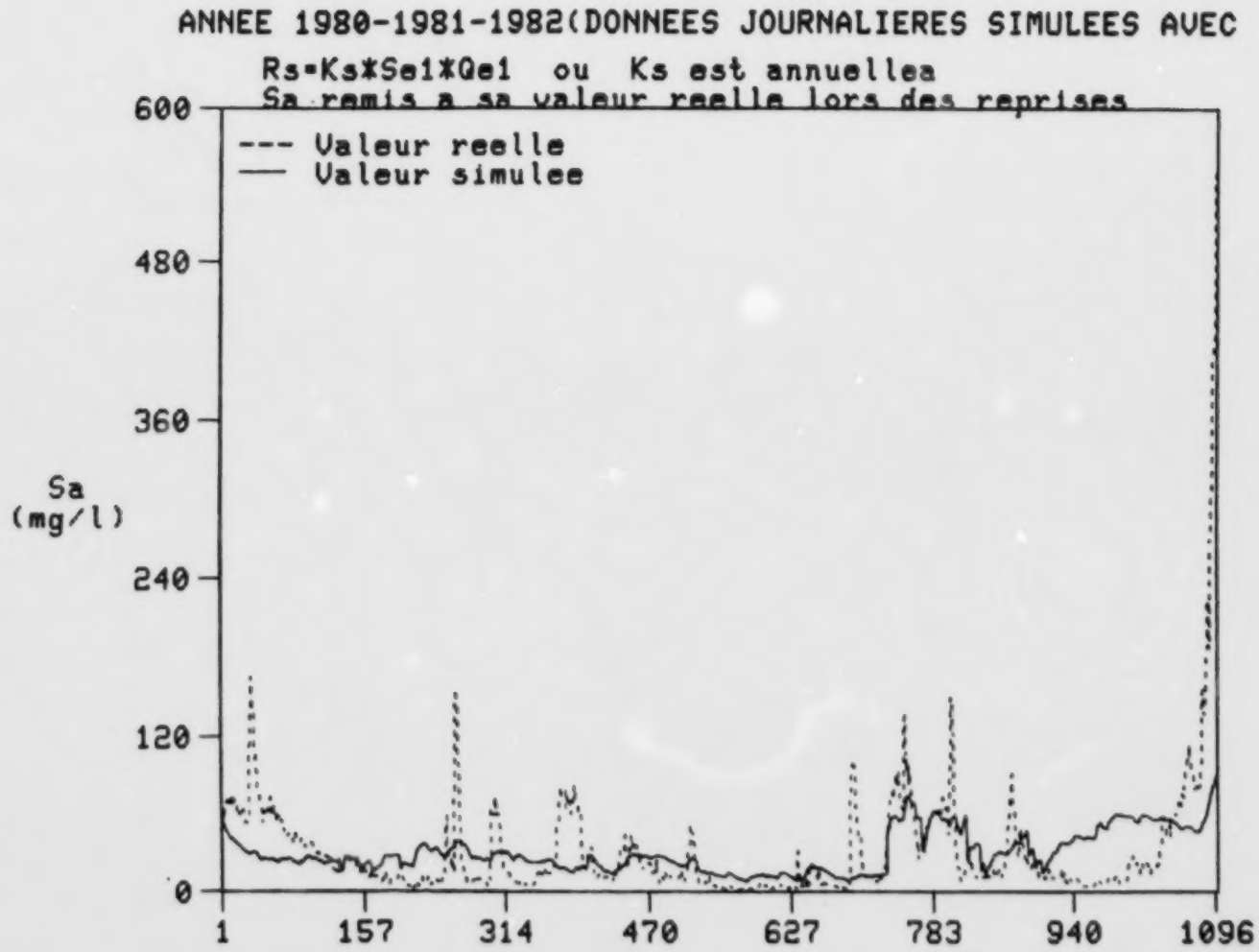

ANNEE 1980-1981-1982(DONNEES JOURNALIERES SIMULEES AVEC
Rs=Ks*Se1*Qe1 ou Ks est annuellea
Sa remis a sa valeur reelle lors des reprises
--- Valeur reelle
—— Valeur simulee
Sa
(mg/l)
600
480
360
240
120
0
1
157
314
470
627
783
940
1096
1980-1981-1982(JOUR)

FIGURE 12

TABLEAU 2: VALEURS ANNUELLES MOYENNES DES PARAMÈTRES $\bar{K}_x$ ET $\bar{K}_x$ UTILISÉS POUR DIVERSES FORMULATIONS

FORMULATION:	$R_s = \bar{K}_s . Q_e . S_e$	$R_x = \bar{K}_x$	$R_x = \bar{K}_x . Q_e . S_e$
Paramètres:	$\bar{K}_s$ (litre^{-1})	$\bar{K}_x$ (mg litre^{-1} min^{-1})	$\bar{K}_x$ (litre^{-1})
Année 1980	0.17146×10^{-6}	0.18×10^{-1}	0.12×10^{-7}
Année 1981	0.17187×10^{-6}	0.37×10^{-1}	0.33×10^{-7}
Année 1982	0.16998×10^{-6}	0.25×10^{-1}	0.25×10^{-7}

(Bellavance, 1983). Toutes les variations temporelles ont une importance relative égale. Cependant, comme il est possible de l'observer à la Figure 13 le paramètre K_s, pour sa part, possède une périodicité importante de 12 mois La présence d'une densité spectrale majeure à 12 mois indique que les valeurs journalières de K_s exibent un cycle annuel. De plus, le tracé de ces valeur de K_s en fonction du temps montre une tendance similaire à celle observée avec la température de la liqueur mixte (Figure 5). Bien qu'en général une dépendance exponentielle en fonction de la température soit rapportée pour des taux de réaction, la comparaison des valeurs de K_s avec celles de la tempéra ture pour la période entière de 3 années a permis d'établir l'adéquation de la relation linéaire suivante:

$$R_x = R_{so} + R_{x1} . T \qquad (13)$$

pour la plage de température considérée. T représente la température (°K) et:

$$R_{so} = 0.14059 * 10^{-6} \text{ (litre}^{-1}\text{)}$$

$$R_{s1} = 0.10626 * 10^{-9} \text{ (litre}^{-1} \text{ °K}^{-1}\text{)}$$

L'utilisation du paramètre $\bar{K}_s$ de la relation (13) dans l'expression $R_s = \bar{K}_s . Q_e . S_e$ produit le résultat illustré à la Figure 14. La comparaison entre la Figure 12 et la Figure 14 montre l'amélioration significative des résultats même si des valeurs constantes sont utilisées sur une période de 3 années de simulation pour R_{so} et R_{s1}.

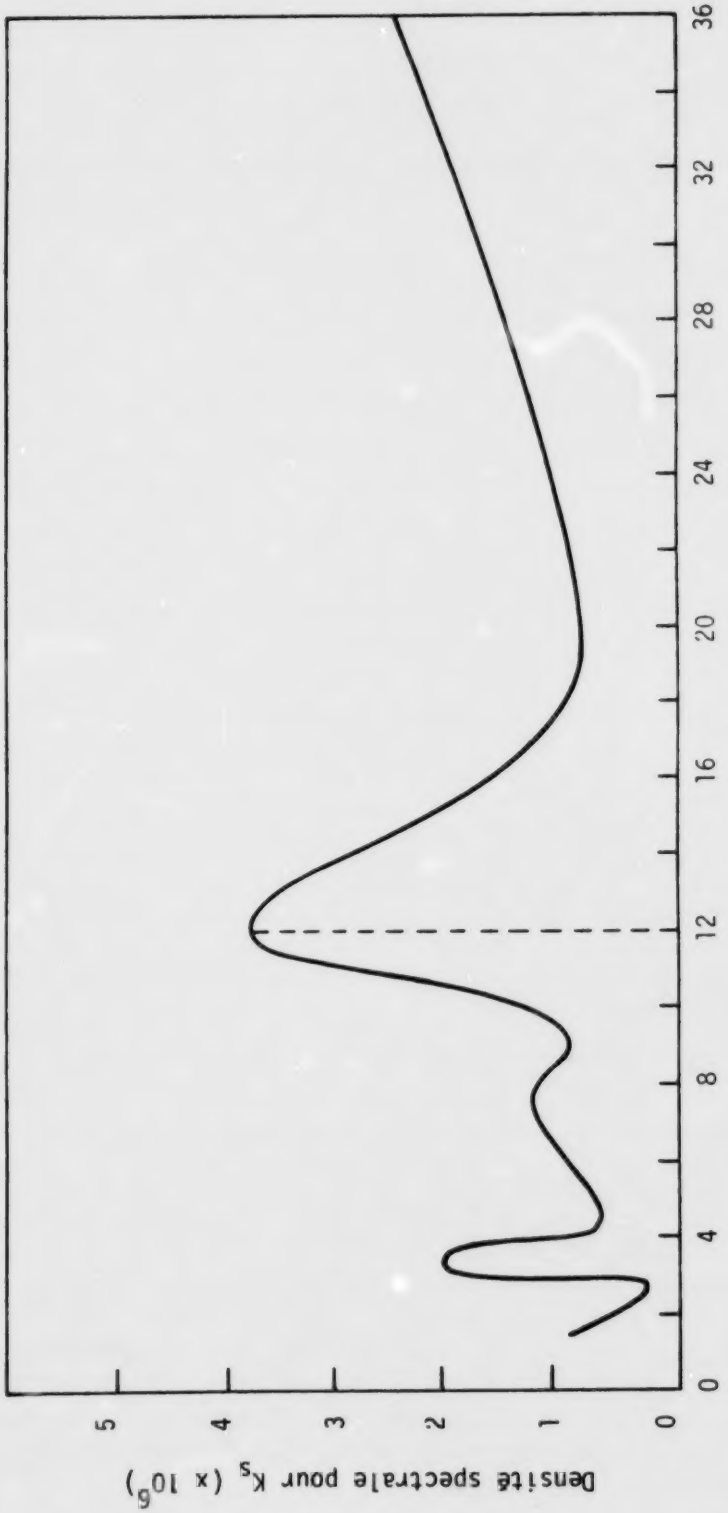


Figure 13: Analyse spectrale

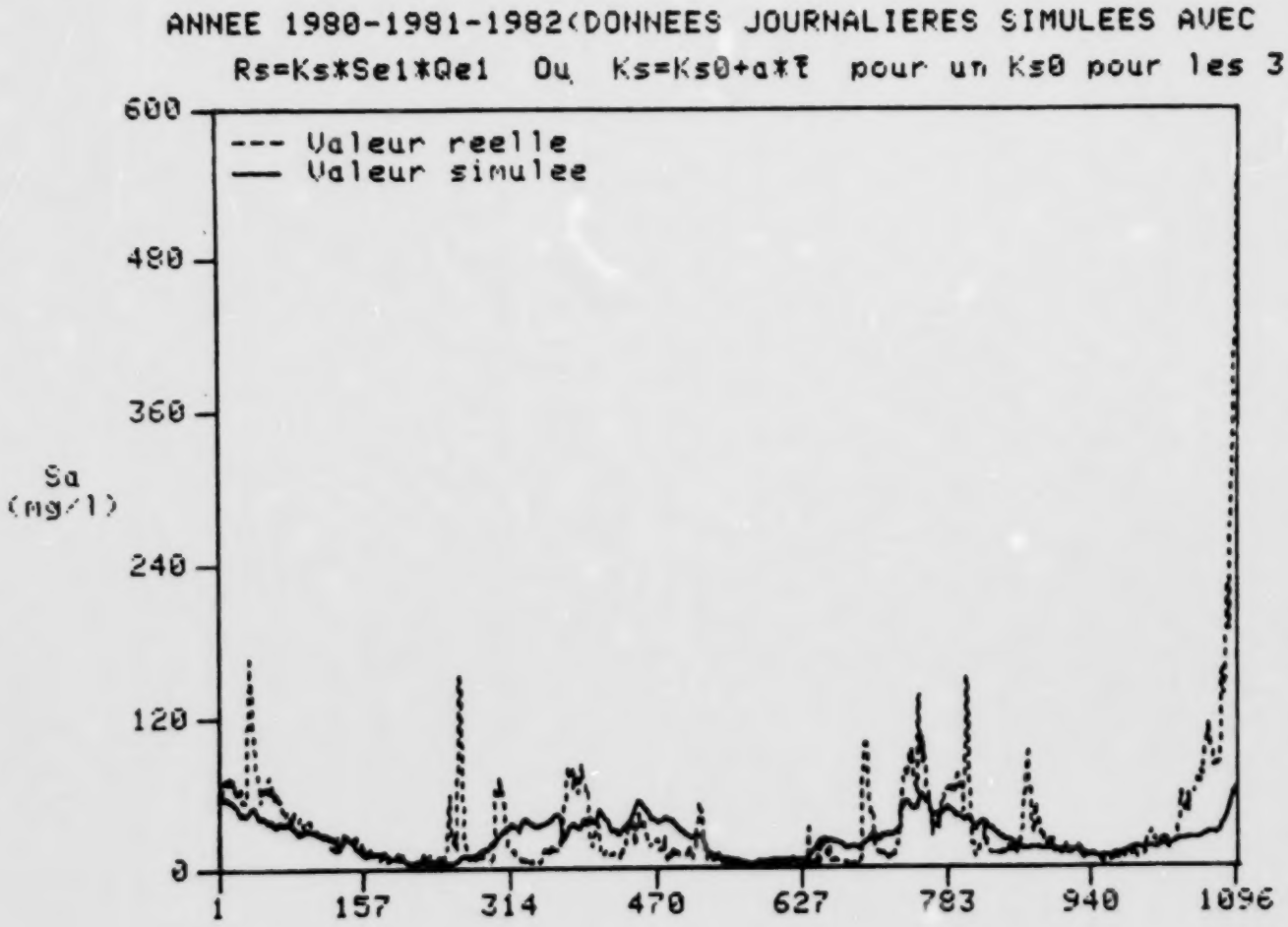

ANNEE 1980-1981-1982(DONNEES JOURNALIERES SIMULEES AVEC
Rs=Ks*Se1*Qe1 Ou Ks=Ks0+a*t pour un Ks0 pour les 3 ans
--- Valeur reelle
—— Valeur simulee
600
480
360
240
120
0
Sa
(mg/l)
1
157
314
470
627
783
940
1096
1980-1981-1982(JOUR)

FIGURE 14

Les travaux en cours étudient les performances relatives de ces formulations pour des paramètres $\bar{K}_x$ évalués sur une période plus courte (trimestrielle, mensuelle, hebdomadaire). Les résultats devraient alors indiquer la période critique à retenir pour la représentation adéquate du comportement dynamique du système et suggérer ainsi les conditions sous lesquelles un tel modèle pourrait être utile pour la prévision à court terme de l'évolution du système et l'élaboration de scénarios d'intervention.

CONCLUSION

Un modèle consistant de deux équations différentielles et décrivant le comportement dynamique du procédé de boues activées a été posé à partir de bilans massiques différentiels sur le substrat et les solides en suspension du système.

Ce modèle retient comme variable significative du procédé le substrat organique biodégradable (exprimé en DBO_5) et la biomasse (exprimée par le contenu en solides en suspension). Il incorpore les variables de perturbations du procédé en considérant explicitement la charge organique de la liqueur à traiter et entrant au procédé. Il inclue également et de façon explicite le paramètre opérationnel qu'est le débit d'extraction des boues du procédé.

Le modèle fut utilisé afin d'extraire des données disponibles, les taux journaliers d'assimilation du substrat et de croissance de la biomasse pour une période de trois ans. Ceux-ci furent comparés à une série d'expressions cinétiques couramment rapportées dans la littérature ainsi qu'avec des formulations limites du modèle de stockage du substrat dans les cellules.

Les résultats obtenus en utilisant des paramètres constants sur des périodes successives d'un an montrèrent que seuls quelques formulations s'avérèrent satisfaisantes.

En particulier, l'utilisation d'une expression pour le taux d'assimilation fonction de la charge organique entrant au procédé et compatible avec le concept de stockage du substrat dans les cellules généra les meilleurs résultats. Le résultat fut significativement amélioré lors de la considération de l'influence du cycle annuel de température sur le paramètre $\bar{K}_s$ de cette formulation. Pour le taux de croissance R_x deux formulations, chacune représentative de conditions extrêmes pouvant exister pour le procédé de boues activées, génèrent des résultats satisfaisants mais jugés complémentaires les uns des autres.

Finalement, le modèle posé s'avère relativement apte à décrire les tendances dynamiques observées pour le système même sous des conditions où les paramètres $\bar{K}_s$ et $\bar{K}_x$ du modèle sont maintenus constants sur une base annuelle. Les travaux en cours indiquent que l'utilisation de paramètres évalués plus fréquemment permet d'obtenir des résultats encore plus satisfaisants pour le modèle.

REMERCIEMENTS

Les auteurs désirent remercier chaleureusement MM G.G. Gagnon et J. Herbineaux, respectivement directeur et assistant technique de l'usine pour

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L'EFFET DE POLYELECTROLYTES UTILISES COMME FLOCULANTS SUR DES MICRO-ORGANISMES PRESENTS DANS LES EAUX RECEPTRICES

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INTRODUCTION

Les polyéléctrolytes (ou polymères synthétiques chargés) sont fréquemment employés dans les procédés de traitement des eaux, eaux usées et boues. Dans les procédés de traitment des boues, les polyéléctrolytes sont ajoutés directement à celles-ci pour en améliorer la floculation. On croyait que les polyéléctrolytes ainsi ajoutés s'adsorberaient complètement sur les flocs présents dans ces boues. Des études effectuées par Gehr et Henry (1982) et par Goppers et Straub (1976) ont démontré que le polyéléctrolyte ne se dépose pas entièrement, mais qu'une concentration résiduaire de ce floculant reste en solution et est conséquemment rejetée dans les eaux traitées qui sont souvent divergées vers une rivière ou tout autre eau naturelle.

Des essais de toxicité sur des truites arc-en-ciel (Salmo gairdneri) et autres organismes aquatiques ont démontré que les polymères synthétiques étaient toxiques à des concentrations aussi faibles que 0,5mg/L (Biesinger et al, 1976; Spraggs, Gehr et Hadjinicolaou, 1982).

En tenant compte des faits cités, des essais sur la toxicité potentielle de ces mêmes polyéléctrolytes sur des micro-organismes aquatiques furent menées et sont présentées dans la présente étude. L'importance de cette éventuelle toxicité est reliée au rôle primordial des bactéries aquatiques dans la purification des eaux naturelles. Il est important de souligner que la biodégradation possible de ces polyéléctrolytes par certaines espèces de microbes aquatiques pourrait influencer soit négativement, soit positivement l'effet toxique de ceux-ci sur la faune aquatique. Par exemple, l'acrylamide, un monomère composant plusieurs polymères couramment utilisés, est reconnu comme étant toxique envers les animaux possédant des systèmes nerveux centraux.

METHODOLOGIE

La présente étude se divise en quatre sections:

A) étude de toxicité des polymères sur des bactéries aquatiques mixtes. Croissance bactérienne contrôlée par mesurage de la densité optique;

B) étude de toxicité des polymères et de monomères sur des bactéries aquatiques mixtes. Croissance indiquée par mesurage de la consommation d'oxygène des bactéries;

C) isolation d'une espèce de bactérie aquatique pouvant avoir la faculté de biodégrader un polymère désigné;

D) étude de biodégradation de plusieurs polymères séléctionnés par l'espèce

bactérienne isolée en C). Croissance mesurée par consommation d'oxygène;

A) Etude de toxicité sur des bactéries aquatiques mixtes.

Cultures bactériennes

Dans le but d'étudier l'effet toxique des polymères sur la faune microbienne aquatique, les expériences furent effectuées avec une population batérienne mixte isolée du fleuve Saint Laurent (Ile des Soeurs, près de Montréal). Les bactéries furent isolées par l'ensemencement de tubes de cultures contenant de l'agar nutritif et portées à l'étuve à 20°C pendant 48 heures. Un tel procédé permet d'isoler uniquemment des bactéries aquatiques non fastidieuses.

Solutions d'essai

Les polymères enrichis de sels minimaux constituaient les solutions d'essai de base. Les deux polymères testés dans cette section, sont un polyacrylamide co-polymère désigné comme Polymère A (pour aider la dissolution du polymère, de l'éthanol fût ajouté à la solution de base) et un homo-polymère de chlorure d'ammoniaque de diallyl diméthylique ci-nommé Polymère C, tous deux couramment employés en floculation. Les solutions de polymères furent enrichies de glucose dans le but d'évaluer la toxicité des polymères sur les bactéries ensemencées. Les solutions d'essai étaient alors formées par le mélange en différentes proportions des solutions de polymères et de glucose (recettes au Tableau 1).

Essai de toxicité

Les solutions d'essai dûment ensemencées avec 1% de cultures bacteriennes mixtes (1mL de bactéries diluées en solution saline dans 100mL de solution d'essai) étaient mesurées à intervalles réguliers pour leur densité optique. Le mesurage était exécuté stérilement à l'aide de flasques de cultures de 250mL à tubes latéraux (de 29mL) insérées dans un spéctrophotomètre Bausch et Lomb (modèle No.20) à 690nm.

Tableau 1. RECETTES DES SOLUTIONS D'ESSAI

Solutions (mg/L)	Sels* (mg/L)	(PC) (mg/L)	(PA)*** (mg/L)	(MC) (mg/L)	(MA) (mg/L)	(PD) (mg/L)	(glu) (mg/L)	(2318)** (mg/L)
gluc(2000)	1210	--	--	--	--	--	2000	--
gluc(1000)	1210	--	--	--	--	--	1000	--
PolC(2000)	1210	2000	--	--	--	--	--	--
PolC(1000)	1210	1000	--	--	--	--	1000	--
MonC(2000)	1210	--	--	2000	--	--	--	--
MonC(1000)	1210	--	--	1000	--	--	1000	--
PolA(2000)	1210	--	2000	--	--	--	--	--
PolA(1000)	1210	--	1000	--	--	--	1000	--
MonA(2000)	1210	--	--	--	2000	--	--	--
MonA(1000)	1210	--	--	--	1000	--	1000	--
PolD(2000)	1210	--	--	--	--	2000	--	--
2318(2000)	1210	--	--	--	--	--	--	2000

Légende;gluc(2000)= sol. de 2000mg/L de glucose; PolC(1000)= sol. de 1000mg/L de Polymère C; MonC= sol. de Monomère C; PolA= sol. de Polymère A; MonA=sol. de Monomère A; 2318= sol. de Pol. 2318.
*Recette de sels:20g chlorure d'ammoniaque;4g nitrate d'ammoniaque; 8g sulfate de sodium;12g phosphate de potassium (monohydrate); 4g phosphate de potassium (dihydrate);0,4g sulfate de magnésium; dilué à 1L.
** exemple des trois polymères 2318, 2319/1 et 2319/4.
*** Dans les études de toxicité, la solution de Pol. A était enrichie d'éthanol.

B) Etude de toxicité des polymères et de leurs monomères par consommation d'oxygène.

Culture bactériennes
Les mêmes cultures que celles décrites précédement furent utilisées ici.

Solutions d'essai
Les polymères expérimentés dans la section précédente furent employés à nouveau ici, mais en plus le Monomère C (monomère de chlorure d'ammoniaque de diallyl diméthylique) et le Monomère A (acrylamide) furent également expérimentés. Les recettes de ces solutions d'essai sont présentées au Tableau 1.

Essai de toxicité
Les solutions de polymère et monomère ensemencées étaient controléees pour toute croissance microbienne par la quantité d'oxygène consommée par les bactéries actives. La consommation d'oxygène était mesurée par un respiromètre éléctrolytique (Exidyne, modèles ER-101 et ER-102) (Larson et Perry, 1981). Un régulateur de température Leedal maintenait celle-ci constante. A intervalles réguliers, la concentration cumulative d'oxygène consommé par chaque réacteur (5 réacteurs opéraient simultanément) était imprimée et tracée. Chaque expérience comprenait 1 réacteur contrôlant l'effet baromètrique (eau distillée), 1 réacteur mesurant la respiration endogène (sels minimaux ensemencés), 1 réacteur

témoin (solution de glucose ensemencée) et 2 réacteurs d'essai (solutions d'essai).

C) Choix de bactéries aquatiques biodégradant les polymères

Cultures bactériennes
Des cultures pures de bactéries aquatiques furent isolées à partir de boîtes de Petri contenant de l'agar nutritif ensemencées avec une goutte d'eau fluviale. Les colonies séléctionnées furent transférées dans de nouvelles boîtes de Petri. Les solutions d'essai furent donc ensemencées avec ces cultures pures obtenues par transfer successif.

Solutions d'essai
La solution de Polymère C (2000mg/L) fût séléctionnée pour ces essais de biodégradation.

Essai de biodégradation
La biodégradation de la solution de Polymère C fût mesurée par l'oxygène consommé par les bactéries et donc par le respiromètre éléctrolytique. Etant donné que le polymère représente la seule source de carbone dans la solution ensemencée, toute consommation significative d'oxygène était interprétée comme une biodégradation du polymère.

D) Etude de biodégradation des polymères

Cultures bactériennes
La culture bactérienne qui consomma une concentration d'oxygène significative dans la section précédente (culture No.5, Figure No.6), fût utilisée dans toutes les expériences ci-après. Les cultures furent maintenues en laboratoire par des transfers bi-hebdomadaires dans de l'agar nutritif frais.

Solutions d'essai
Les solutions de polymères testées ici, représentent un assortiment de polymères anioniques et cationiques utilisés dans les procédés de traitement des eaux, eaux usées et boues. Les recettes de ces solutions sont décrites au Tableau 1. Etant donné que les solutions étaient uniquement testées pour leur biodégradation, aucune quantité de glucose ne fût rajoutée comme dans les sections A) et B). Les polymères soumis à l'essai étaient: Le Polymère A, le Polymère C (dialysé pour le purifier) le Polymère D (un polyacrylamide anionique), le polymère 2318 (un polymère à base d'acrylamide avec 50% d'acrylate éthylique amino diméthylique quaternisé avec du chlorure de méthyle), et les polymères 2319/4 et 2319/1, tous deux similaires au 2318 mais comprenant uniquement 20% d'acrylate.

Essais de biodégradation
La biodégradation des divers polymères ensemencés avec le Pseudomonas isolé, fût mesurée par la consommation d'oxygène.

Essai d'analyse des nutrients résiduels
Dans le but d'affirmer l'assimilation des polymères par le Pseudomonas, des essais d'analyse du Polymère C, Monomère C et glucose non-consommés après incubation, furent menés en analysant la solution d'essai finale par la méthode

de DCO telle que décrite dans les méthodes normalisées américaines (APHA et al, 1980). Cependant, pour éviter d'analyser les bactéries, les polymères restant en solution et ceux adsorbés aux bactéries ont été séparés par l'ajout de 1.0g/100mL de chlorure de sodium à la solution. Ceci entraîne une inversion des charges ioniques et libère les polymères. Les solutions de polymère, de monomère et de glucose furent ensuite centrifugées à 10 000rpm.

RESULTATS

L'utilisation de moyens optiques pour mesurer l'activité métabolique microbienne a permis de tracer des courbes de croissance semblables à celles développées à partir des données sur la consommation d'oxygène des mêmes bactéries. Ainsi, on peut constater que ni le Polymère C, ni le Polymère A ne supportaient autant de croissance microbienne qu'une concentration égale de glucose (Figures 1 (courbes 2, 3 et 4), 4 (courbe 5) et 5 (courbe 5)). Cependant, la solution de 2000mg/L de Polymère C ne supportait aucune croissance ou consommation d'oxygène supérieure à la respiration endogène (obtenue avec les sels minimaux) (Figures 1(courbes 1 et 2) et 5 (courbes 2 et 3)). La solution de Polymère A (2000mg/L) a provoqué une certaine activité métabolique qui demeure cependant inférieure à celle causée par le glucose (Figures 1 (courbe 3) et 4 (courbe 3)) et peut être tentativement expliquée par la présence d'éthanol et d'impuretés dans la solution d'essai.

Cependant, quand les solutions de polymères étaient enrichies de glucose, les micro-organismes répondaient par un accroissement de leur activité métabolique en comparaison avec la courbe témoin (glucose) (voir Figures 2 (courbes 1, 2 et 3), 4 (courbe 5) et 5 (courbe 4)).

Des essais de toxicité ont également été conduits avec deux monomères: Le monomère de chlorure d'ammoniaque de diallyl diméthylique (appellé Monomère C), seul constituant du Polymère C, et l'acrylamide, constituant de base du Polymère A, Polymères 2319/1, 2319/4 et 2318 et du Polymère D. Ces monomères se sont également montrés toxiques envers les truites arc-en-ciel lorsqe soumis à essais de toxicité aquatiques et dynamiques (Spraggs et al, 1982). Les résultats obtenus avec les monomères sont, en tout point de vue, comparables avec les résultats obtenus avec leur polymère correspondant (Figures 4 (courbes 2 et 4) et 5 (courbes 1 et 5)). Cependant, en l'absence de glucose, le Monomère A n'a pu soutenir une activité métabolique comme le Polymère A (Figure 4 (courbe 2)), fait d'ailleurs explicable par l'absence d'éthanol dans la solution de monomère ainsi que son degré de pureté supérieur.

Vu que les polymères testés n'ont point démontré de toxicité vis-à-vis la majorité des microbes ensemencés, et qu'en la présence de glucose un effet activateur s'est fait remarquer, il a été décidé de poursuivre cette étude vers la biodégradation possible des polymères. Pour atteindre ce but, il fallut tout d'abord trouver une espèce microbienne aquatique ayant la faculté d'assimiler les polyéléctrolytes. Les résultats de cette recherche sont présentés à la Figure 6.

L'espèce bactérienne démontrant la faculté de biodégrader le Polymère C (culture No.5), se caractérise par sa forme de batonnet, Gram négatif, colonies beiges et odorantes. La bactérie a par ailleurs été identifiée comme faisant partie du genre Pseudomonas, mais une identification de l'espèce de put être accomplie.

Le Pseudomomas isolé du fleuve fût ensemencé dans une variété de

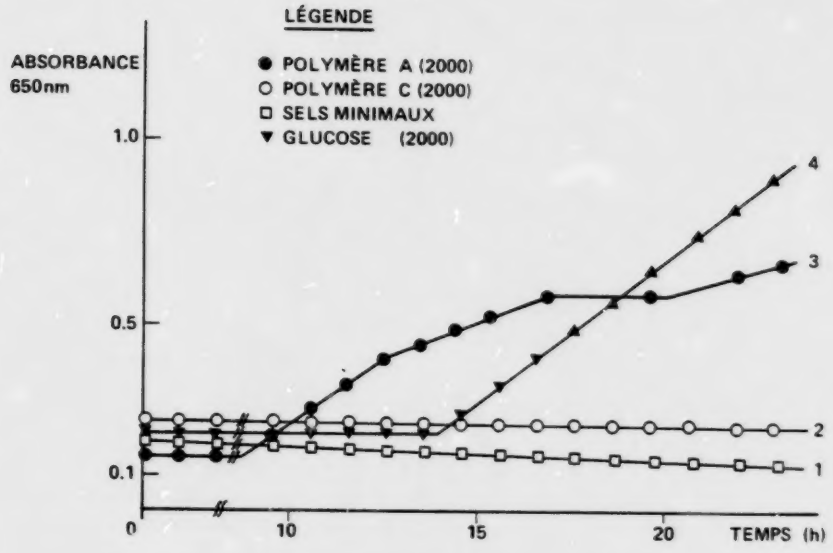

Figure 1. COURBES DE CROISSANCE DE MICROORGANISMES AQUATIQUES MIXTES ENSEMENCÉS DANS LES SOLUTIONS D'ESSAI ET INCUBÉS À 14°C.

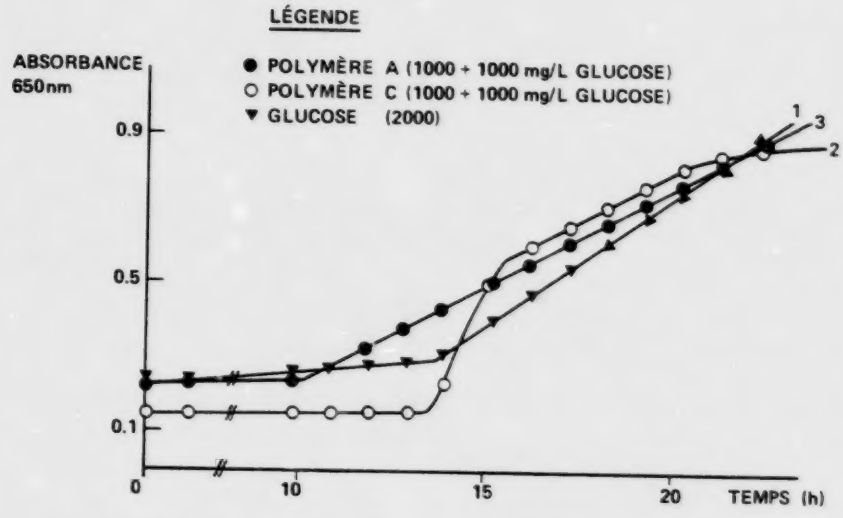

Figure 2. COURBES DE CROISSANCE DE MICROORGANISMES AQUATIQUES MIXTES ENSEMENCÉS DANS LES SOLUTIONS D'ESSAI ENRICHIES ET INCUBÉS À 14°C.

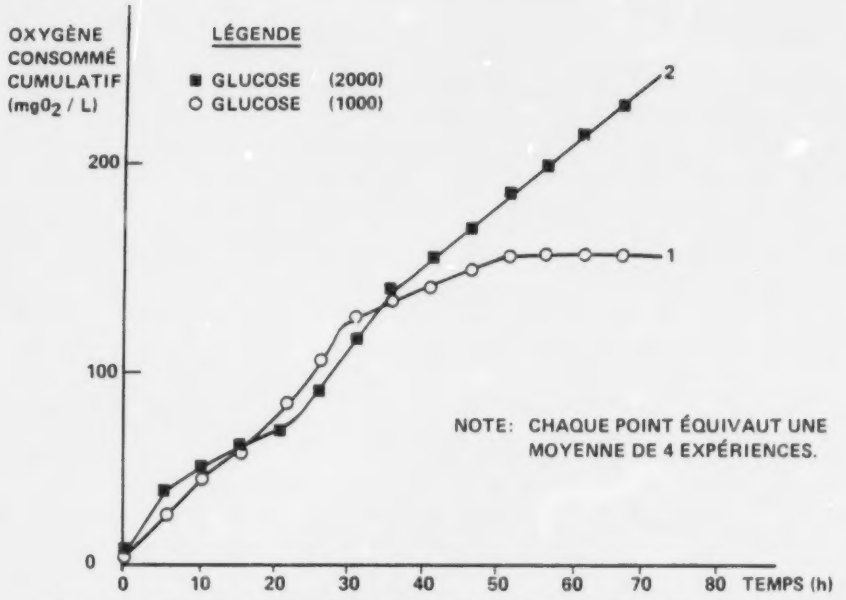

Figure 3. COURBES DE CONSOMMATION D'OXYGÈNE PAR DES MICROORGANISMES AQUATIQUES MIXTES ENSEMENCÉS DANS DES SOLUTIONS DE GLUCOSE ET INCUBÉS À 14°C.

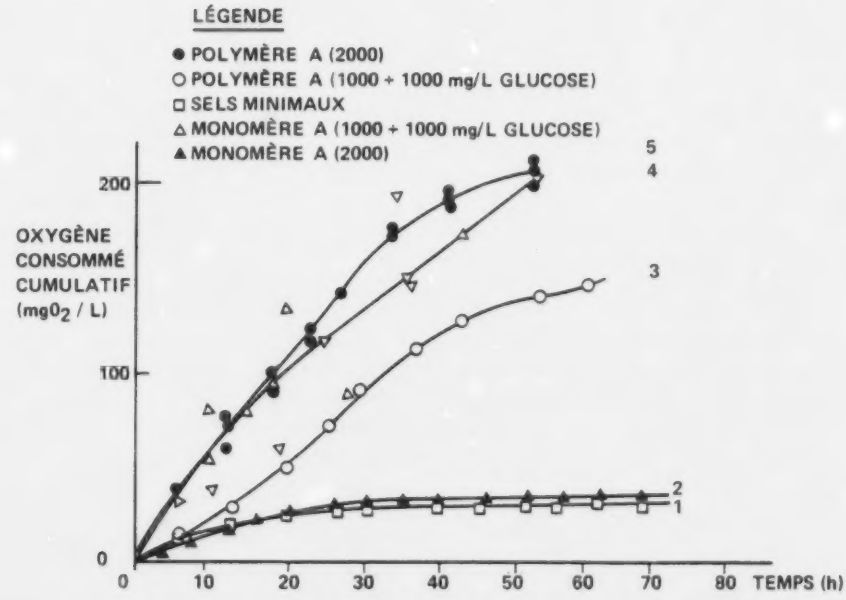

Figure 4. COURBES DE CONSOMMATION D'OXYGÈNE PAR DES MICROORGANISMES AQUATIQUES ENSEMENCÉS DANS DIVERSES SOLUTIONS D'ESSAI ET INCUBÉS À 14°C.

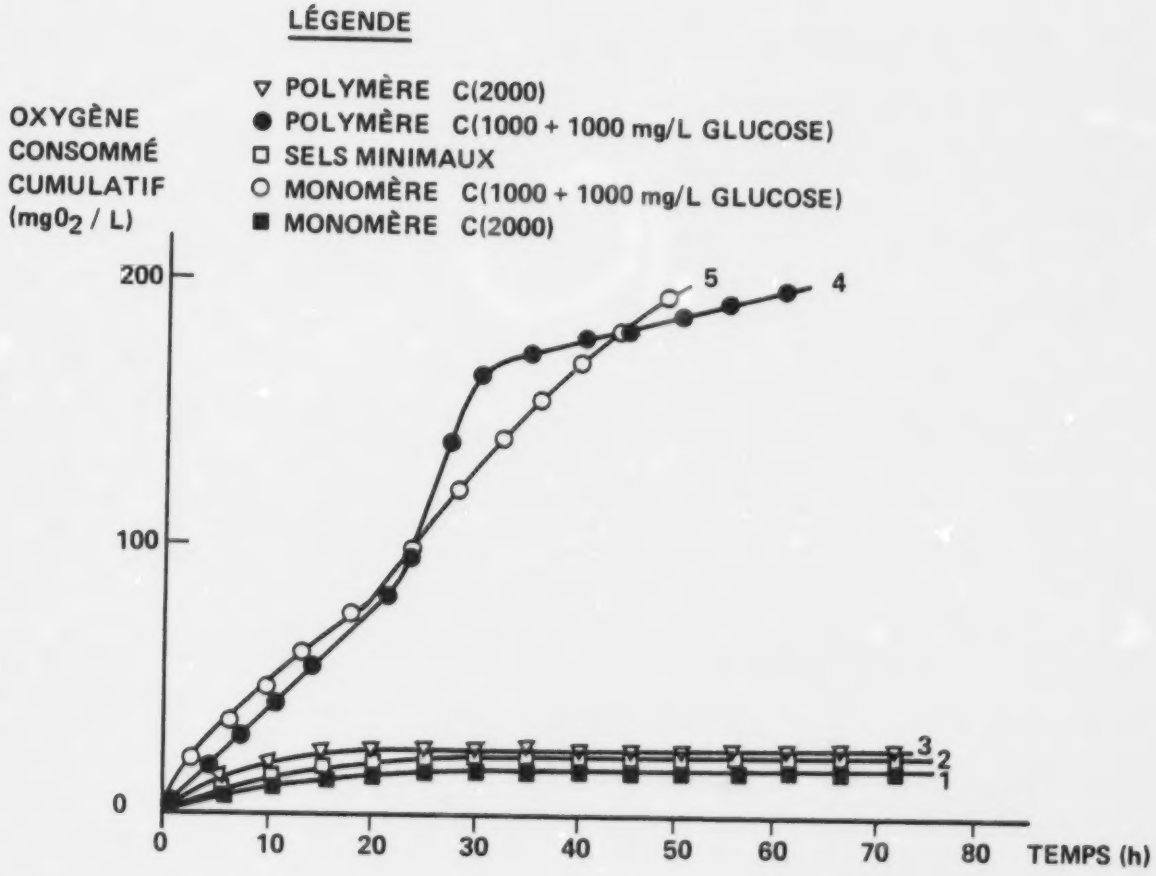

Figure 5. COURBES DE CONSOMMATION D'OXYGÈNE PAR DES MICROORGANISMES AQUATIQUES MIXTES ENSEMENCÉS DANS DIVERSES SOLUTIONS D'ESSAI ET INCUBÉS À 14°C.

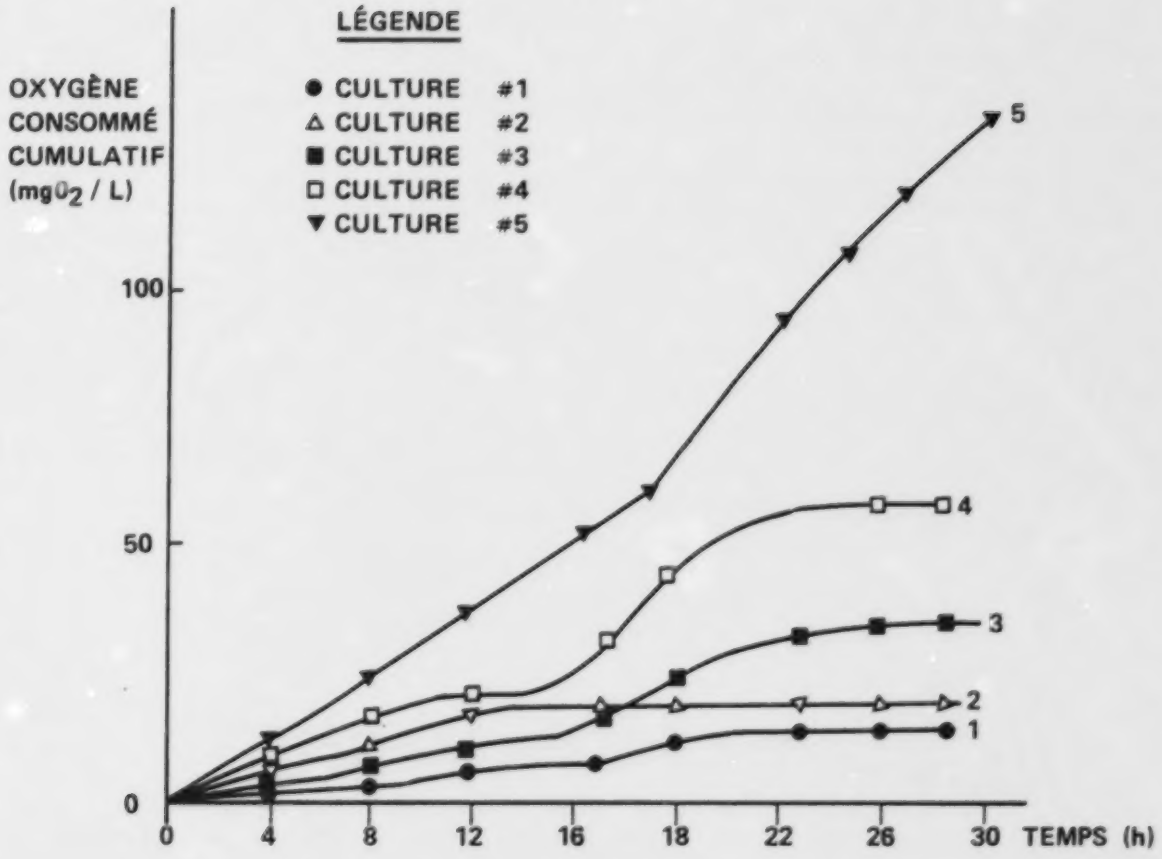

Figure 6. COURBES DE CONSOMMATION D'OXYGÈNE PAR DES CULTURES PURES DE MICROORGANISMES AQUATIQUES ENSEMENCÉS DANS UNE SOLUTION DE POLYMÈRE C ET INCUBÉS À 14°C.

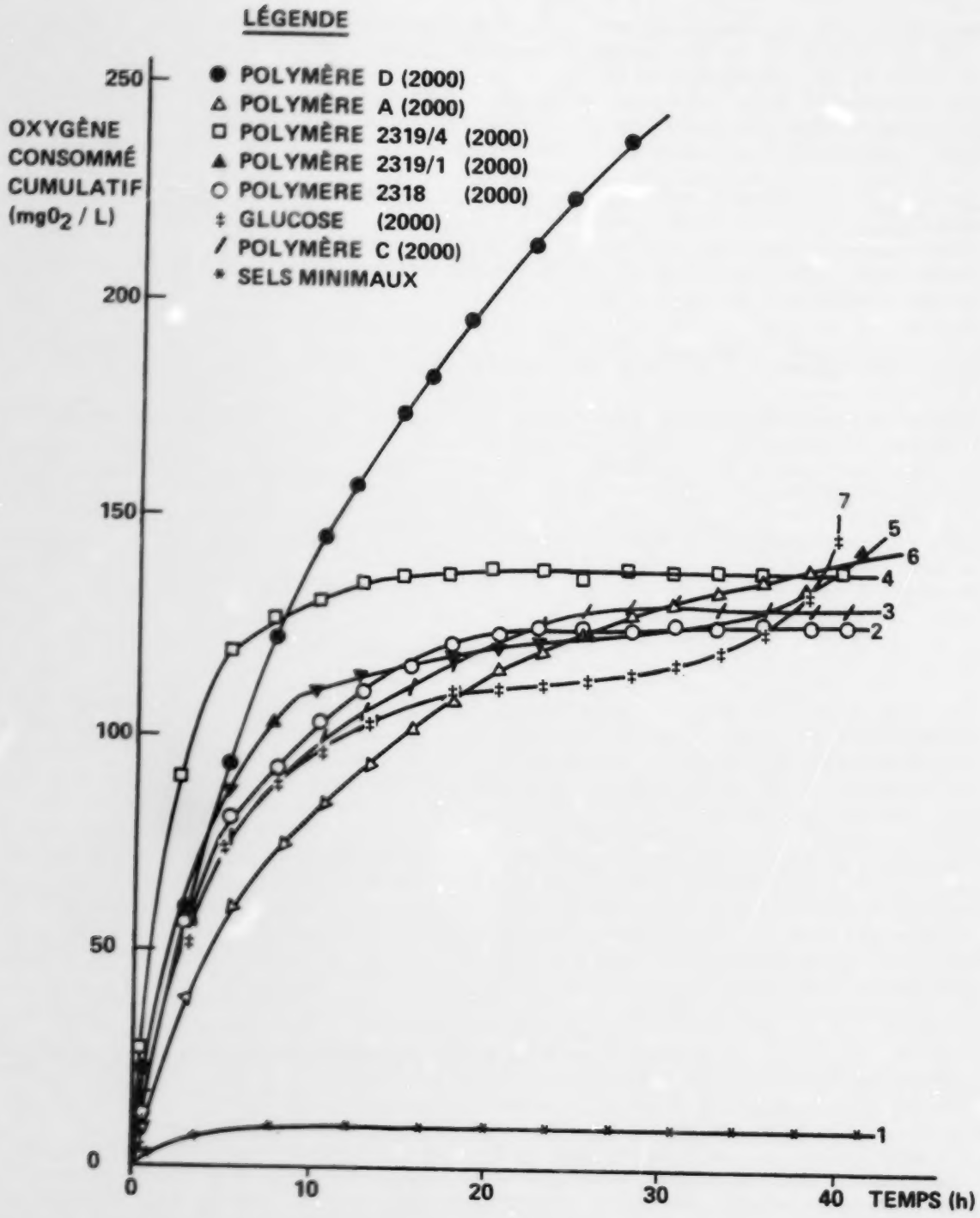

Figure 7. COURBES DE CONSOMMATION D'OXYGÈNE PAR DES CULTURES PURES D'UNE ESPÈCE AQUATIQUE DE PSEUDOMONAS ENSEMENCÉES DANS DES SOLUTIONS VARIÉES ET INCUBÉES À 20°C.

polyéléctrolytes couramment utilisés comme floculants. Tous les polymères testés ont pu servir de seule source de carbone, supportant une activité métabolique intense de la bactérie, qui se caractérise par une nette consommation d'oxygène et croissance (celle-ci était vérifiée par la différence de turbidité entre les solutions d'essai initiales et finales). Le Pseudomonas sembla donner préférence au Polymère D qui diffère des autres par sa charge anionique.

Pour confirmer que la consommation d'oxygène observée dans les essais présentés peut être correlée à une biodégradation du polymère, des analyses de nutrients résiduels furent menées sur le glucose, le Monomère C et son Polymère. Les résultats obtenus (Tableau 2) démontrent qu'effectivement la DCO dans la substance nutritive régresse, avec une diminution égale à 34% après 20 heures d'incubation avec le Pseudomonas isolé.

Tableau 2. RESULTATS DES ANALYSES DE NUTRIENTS RESIDUELS.

Solution d'essai	Concentration initiale (mg/L)	Concentration initiale en DCO (mg/L)	Concentration finale en DCO (mg/L)	Diminution de la DCO (%)
Pol. C	2000	650	400	38,5
Mon. C	2000	800	544	32,0
Gluc.	2000	675	435	32,9

DISCUSSION DES RESULTATS

Les résultats obtenus avec le Polymère A , le Polymère C, le Monomère A et le Monomère C, permettent de conclure que ni les polymères ni leurs monomères respectifs n'inhibent la croissance, ou la consommation d'oxygène, des bactéries aquatiques mixtes ensemencées. Cette conclusion est basée sur les résultats obtenus avec les solutions d'essai enrichies de glucose (Figures 2 (courbes 2 et 3), 4 (courbes 4 et 5) et 5 (courbes 4 et 5)). En résumé, les polymères et monomères présents en concentrations aussi elevées que 1000mg/L n'empêchaient pas certaines ou toutes les bactéries ensemencées d'oxider le glucose et donc de consommer de l'oxygène et croître. L'ensemencement de cultures mixtes pour des essais de toxicité représente une situation plus naturelle que l'ensemencement de cultures pures, mais ne permet pas de discuter sur la toxicité ou l'absence de toxicité des polymères sur des espèces bactériennes en particulier.

Quoique les polymères et les monomères testés (solutions de 2000mg/l) (à l'exclusion du Polymère A) n'ont pas démontré de signe de toxicité, ils n'ont pas pour autant supporté la croissance de ces mêmes bactéries, ne pouvant pas servir de seule source de carbone (Figures 1 (courbe 2), 4 (courbe 2) et 5 (courbes 1 et 3)). Cependant, quand ces mêmes solutions d'essai étaient pourvues d'une source de carbone facilement oxidable tel le glucose, aussi bien la croissance que la consommation d'oxygène des bactéries étaient activées en comparaison avec le témoin. Donc, sans pourvoir la source de carbone principale, les polymères ainsi que leurs monomères activaient la croissance, probablement en colaborant avec des atomes de carbone et d'azote supplémentaires. Etant donné que les polymères pouvaient activer le métabolisme microbien, ceci peut signifier qu'une dégradation totale ou partielle de la molécule se produit,

libérant des atomes nutritifs. Cette hypothèse est soutenue par les résultats présentés aux Figures 6 et 7 (toutes les courbes).

Une espèce bactérienne aquatique fût, en effet, isolée du fleuve Saint-Laurent. Elle possède la faculté de biodégrader et d'assimiler tous les polymères testés (Figure 7). Un tel pouvoir de biodégradation est d'ailleurs reconnu chez les Pseudomonas qui sont classifiés comme des organismes très adaptables, oxidant des molécules de sulfonates (Clarke et Richmond, 1975; Focht et Williams, 1970; Johnston et al, 1975) de polyéthylenes (Haines et Alexander, 1975), d'hydrocarbures (Van de Linden et Thysse, 1965) et des pesticides (Chatterjee et al, 1982). Cette espèce de Pseudomonas a d'ailleurs consommé plus d'oxygène quand ensemencée dans les polymères plutôt que dans le glucose. Ceci-ci peut s'expliquer par le fait que la bactérie est membre du genre Pseudomonas et que celles-ci ne possédent pas le cycle d'oxidation des sucres par la voie Embden-Meyeroff. L'espèce isolée possède cependant le cycle de Entner-Doudoroff puisqu'à maintes reprises la bactérie a prouvé son habilité à dégrader le glucose.

La présence du Pseudomonas dans les échantillons de cultures mixtes testés auparavant, explique l'activité métabolique rencontrée lorsque les cultures mixtes étaient ensemencées dans les solutions d'essai enrichies de glucose (Figures 2 (courbes 2 et 3), 4 (courbes 4 et 5) et 5 (courbes 4 et 5)). Cette activation peut s'expliquer par la biodégradation du polymère sous l'influence du Pseudomonas, libérant ainsi des atomes de carbone et d'azote disponibles pour les autres bactéries.

CONCLUSIONS

Les essais de toxicité de deux polyéléctrolytes utilisés à titre de floculants sur des cultures bactériennes aquatiques mixtes, ont démontré que ni la consommation d'oxygène ni la croissance totale des cultures étaient affectées par la présence de polymères à des concentrations aussi élevées que 1000mg/L. Cette conclusion, cependant ne specifie pas si des espèces en particulier étaient affectées par ces substances. En d'autres mots, la présence de polyéléctrolytes dans les eaux naturelles n'inhiberait pas l'activité globale des microbes aquatiques et donc ne devrait pas nuire à l'auto-épuration des eaux et à d'autres activités microbiennes.

D'autres études ont cependant démontré que ces mêmes polyéléctrolytes se sont montrés fortement toxiques envers certains organismes aquatiques (Spraggs et al, 1982; Biesinger et al, 1976). La présence de ces polymères dans les eaux est donc fortement déconseillée. Néanmoins, si les polymères sont activement biodégradés par certains micro-organismes aquatiques, leur effet toxique pourrait en être modifié. Les résultats obtenus sur la participation active du Pseudomonas aquatique dans la biodégradation des divers polyéléctrolytes peut laisser planer un certain espoir en ce qui concerne l'inhibition de la toxicité sur la faune aquatique des polymères déversés dans les eaux naturelles. Il est présentement impossible cependant de discuter sur l'inactivation de la toxicité des polymères par leur biodégradation, puisque les sous-produits de biodégradation n'ont pu être testés par des essais de toxicité statiques ou dynamiques sur des macro-organismes aquatiques.

Une meilleure compréhension de la destinée et de l'impact réel des polyéléctrolytes déversés dans les eaux naturelles impliquerait des études plus approfondies. Telles que: Une étude approfondie du procédé de biodégradation des

polyéléctrolytes. Des études de toxicité des sous-produits de biodégradation. Et une connaissance plus détaillée de(s) microbe(s) aquatique(s) ayant la faculté de biodégrader de telles molécules.

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Comparison of Stirred Suspension and Fixed Film Anaerobic Digestion Systems

by

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ABSTRACT

The techno-economic feasibility of two-stage anaerobic digestion systems for the treatment of cellulosic wastes was investigated. In the first stage, the enzymatic hydrolysis of insoluble substrates was the rate limiting step. Optimum cellulose hydrolysis and conversion to volatile acid products was observed at carbon to nitrogen ratio of 24:1. Fixed film and stirred tank reactors were utilized to study methanogenesis (second stage). The kinetics of volatile fatty acid utilization was proven to be rate limiting. Acetate utilization by methanogens followed a Monod-type kinetic model; the maximum specific growth rate (μ_{max}) and the half saturation constant (K_S) were found to be 0.17 day^{-1} and 8.46 x 10^{-3} M acetate, respectively. Propionate utilization, however, was strongly inhibited by acetate. The inhibition resembled classical competitive inhibition patterns, the inhibition constant (K_I) being 2.97 x $^{-7}$ M. The models derived for volatile acid substrates were applicable for acid mixtures in both stirred tank and downflow fixed film reactors.

The economic feasibility of the anaerobic digestion systems has been performed concurrently with the experimental programme. In the analysis, capital equipment costs were based on quotations from manufacturers. Scale-up was based on procedures generally used by the food and fermentation industry. Process economics favoured fixed film biological reactors. A particularly attractive scenario was the co-digestion of paper pulpmill clarifier sludges with animal wastes. Using a carbon to nitrogen ratio of 24:1 in the feed, a fuel gas containing 72% methane was predicted and the process would be economically feasible at moderate installation sizes.

Introduction

Over the past few years anaerobic digestion of organic mater has been regarded as an important waste management option for agricultural, industrial and municipal wastes. Anaerobic digestion provides a potential reduction of pollution in terms of COD or BOD, methane production and recovery of residues as fertilizer. However, its application for the treatment of agricultural wastes has not been widespread due to the long residence time which imply large reactor volume, high initial investment cost, the poor understanding of the process where the optimum conditions are rarely achieved, and frequent digester failure.

Traditionally, anaerobic reactors have been completely mixed free suspension reactors with long residence time. Current research favours

fixed biological film reactors where the biomass is retained on an inert support (Colleran, et al., 1982).

Currently, methane formation is viewed as a three-phase process (Taylor, 1982). The first phase involves acidogenic bacteria which ferment carbohydrates, proteins and lipids to alcohols, volatile fatty acids, H_2 and CO_2. The second phase involves acetogenic bacteria which produce acetate, CO_2 and H_2 from the alcohols and the volatile fatty acids produced in the first stage. Finally, in the third phase, methanogenic bacteria utilizes the products of the previous phases, mainly acetate, CO_2 and H_2, to produce CH_4 and CO_2. Due to the different nutritional and environmental requirements of the microbial groups, stage separation seems to be a rational approach in order to optimize the process (Ghosh and Klass, 1978). The two phases include an acidogenic phase and methanogenic phase.

Research performed in our laboratory has shown that the rate limiting step in the first stage was the enzymatic hydrolysis of insoluble substrates, particularly cellulose (Devlesaver, 1983). It has been observed that the optimum cellulose hydrolysis and conversion to acid products occur at carbon to nitrogen ratio of approximately 24:1. The kinetics of volatile acid utilization (especially acetate and propionate) have been studied in both fixed film and stirred tank reactors. The latter comprises the second stage of a two-stage process. These acids have been chosen for several reasons. Acetate is a major intermediate and can be utilized directly to produce methane. Propionate cannot be converted directly to methane. Since propionate has been shown to accumulate during digester failure, the need for the dynamic characterization of this volatile acid is obvious.

The objectives of this paper are two-fold. First, kinetic models for acetate and propionate utilization are established. Model predictions are compared with our work as well as previously published data. Second, a preliminary economical evaluation is performed for comparing conventional reactors with two stage digestion systems.

Materials and Methods

I. Free Suspension

Inoculum for the kinetic studies was derived from a 20 L methanogenic fermenter, fed with a fatty acid, urea, and salt solution, and operated continuously over a period of 8 months. The cells were separated by centrifugation and resuspended in 300 ml solution (Table I). The pH was adjusted initially at 7.0. The apparatus consisted of two 500 ml erlenmeyer flasks connected with tygon tubing. The first flask contained the 300 ml enrichment culture. In the second flask, 50 ml of 0.1 N Ba $(OH)_2$ was placed for the measurement of CO_2 evolution (Vogel, 1962). The flasks were incubated at 39°C and mixed periodically. Volatile fatty acids, pH, and CO_2 were monitored daily.

II. Fixed Film Bed Reactor

The fixed film bed reactor was a lab scale unit, as shown in figure 1, operated in a downflow manner. The reactor consisted of a glass column of 85 cm height and 6.5 cm inside diameter. The columns were packed randomly with 1.27 cm diameter, ceramic rashing rings to a height of 62 cm. The liquid level was maintained 10 cm above the film support as mentioned by

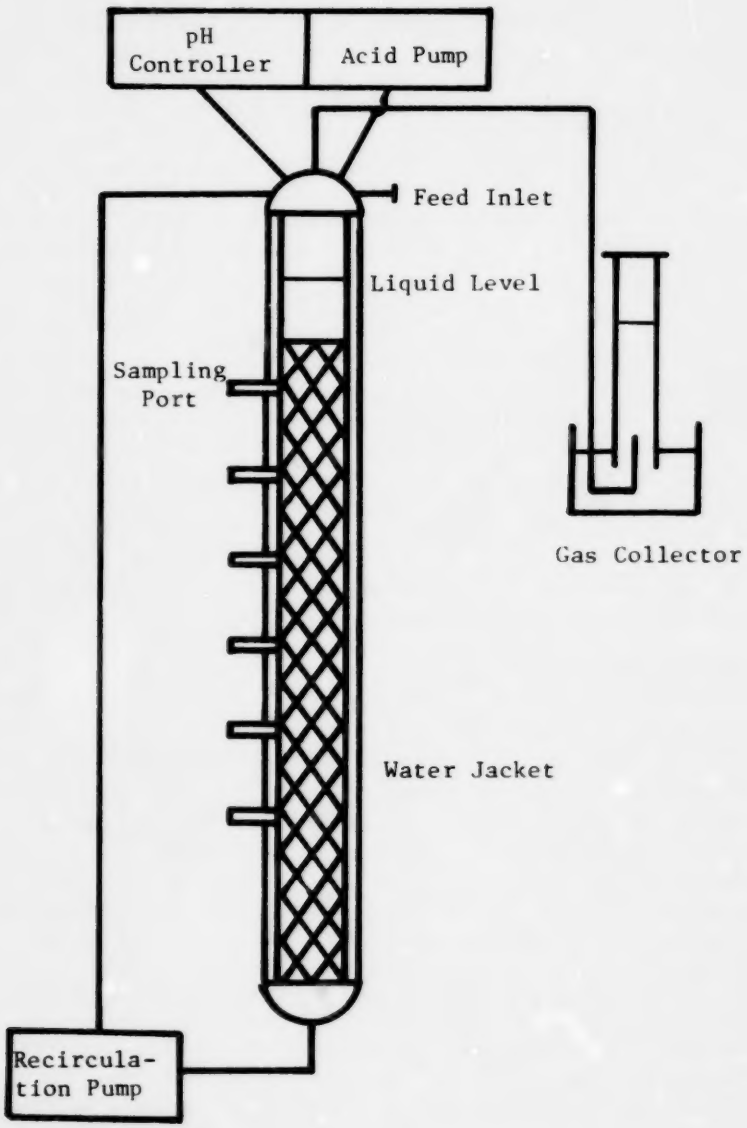


FIGURE 1. SCHEMATIC DIAGRAM OF THE FIXED FILM BED REACTOR.

TABLE I COMPOSITION OF ENRICHMENT MEDIUM FOR ACIDS UTILIZATION

Initial pH	7.00
Acetate	0.03500 and 0.01750 M
Propionate	0.00671 and 0.01242 M
Urea	0.33 g/L
Yeast Extract	0.50 g/L
Salts Solutions	100 ml/L

Salts Solutions Composition based on Natural Swine Manure (Lapp, 1977)

Phosphorus	2.610 g P/L
Potassium	3.850 g K/L
Calcium	3.380 g Ca/L
Aluminium	0.063 g Al/L
Copper	0.017 g Cu/L
Iron	0.145 g Fe/L
Magnesium	0.769 g Mg/L
Manganese	0.035 g Mn/L
Sodium	0.769 g Na/L
Sulfur	0.769 g S/L

Van Den Berg and Kennedy (1982). The surface area to volume ratio was 243 m^2/m^3. The reactor temperature was maintained at 39°C ± 1°C by a water jacket. The liquid volume was recirculated approximately six times per hour with a peristatic pump.

The composition of the enrichment medium is given in Table I. The pH was initially adjusted at 7 with 1N NaOH and was controlled by pH controller and acid addition (HCl, 2N). The inoculum was taken from the same fermentor as for the free suspensions experiments. Three to four months were necessary to build a satisfactory film. Volatile fatty acids, pH, CO_2 and CH_4 were monitored routinely. The gas production was measured by water displacement in inverted graduated flask.

The gas production was analyzed by gas chromatography (Hewlett-Packard Model F & M Scientific 700, USA) using thermal conductivity detector (TCD). The column was packed with Poropak QS 80/100 mesh. The conditions for gas chromatography were as follows: oven temperature = 40°C, flow rate of carrier gas (H_2) = 35 ml/min, temperatures of injection port and detector = 100°C. The volatile fatty acids were analyzed by gas chromatography (Hewlett-Packard Model 5880 A, USA) using flame ionization detector (FID). The column was packed with a 5% NPGSB + 1% H_3PO_4 Anakron A90/100 mesh. The conditions for gas chromatography were as follows: oven temperature = 110°C, flow rate of carrier gas (N_2) = 40 ml/min, temperatures of injection port and detector = 245°C.

Parameter Estimation

Acetate and propionate utilization was modelled by non-linear differential equations. The differential equations were numerically integrated by using the Adams-Moulton algorithm (Henrici, 1962). The subroutine NLDEQD from the University of Waterloo (1965) was used for the numerical integration. The parameters were estimated by minimizing the total sum of squares between observed and predicted acid concentrations by combining the Gauss (Taylor series) method and the method of steepest descent using Marquardt's algorithm (1963). The calculation procedure was iterative. First, the integration was performed with assumed parameters values. The initial parameter values were obtained by varying one parameter of the model at the time and by selecting the value where the minimum total sum of squares was obtained. Then the parameter values were re-evaluated by the Marquardt algorithm and the integration repeated.

The approach based on initial parameter estimates was necessary because of the high correlation between the parameters. In fact, with the non-linear regression method, the same minimum residual sum of squares could be obtained with widely differing parameter values. Some of the parameters values did not correspond with reality. Although they may be acceptable for design purposes, they do not represent a realistic biological system.

The criterion for model fitting were the convergence of the total sum of squares to less than 10^{-5} and the parameter convergence to less than 10^{-5} between two successive iterations. The overall approach is original inasmuch as no previous application of the two algorithms together can be found in the literature.

Results and Discussion

Previous anaerobic reactor designs were based on empirical models (Hashimoto, 1983, Morris et al., 1977, Singh, 1977) which correlate the experimental data on gas production with some loading variable. They do not take into account the mechanism of the anaerobic process and, therefore, cannot predict gas production outside the realm of experimental conditions or digester failure. Mathematical models representing steady state conditions (Scharer, et al., 1982, Ghosh and Pohland, 1974, Lawrence and McCarty, 1969) are more fundamental than empirical models since they attempt to explain methane production from some intermediates present during the fermentation. However, anaerobic reactors are seldom at steady state. For this reason, the dynamic characterization of the process seems the best approach in order to understand, quantify and hopefully predict methane

production. The dynamic models of Andrews and Graef (1971) and the modification by Hill and Barth (1977) have only considered acetate as major intermediate for methane production. However, this approach is an oversimplificiation of the intermediates present during the digestion. Acetate is seldom if ever the only volatile fatty acid present. Higher volatile fatty acids such as propionate are also present. These higher volatile fatty acids are not used directly for methane production, but are degraded by acetogenic bacteria in syntrophic association with methanogens (Taylor, 1982). The presence of high concentration of propionate during digester failure indicates a need for the investigation of these higher acids.

Acetate Utilization

The rate of acetate utilization can be expressed using a Monod kinetic model as follows:

$$\frac{dA}{dt} = \frac{-\mu_{max} AX}{(K_{sa} + A)Y_g} \qquad (1)$$

The concentration of methane bacteria can be related to acetate concentration through the growth yield as it had been done by Ghosh and Klass (1978):

$$\frac{dA}{dt} = \frac{-\mu_{max} A\left[(A_o - A) + \frac{Xo}{Yg}\right]}{(K_{sa} + A)} \qquad (2)$$

The main advantage of equation (2) is that the biomass production is replaced by a more readily measurable term; i.e. acetate concentration. Model parameters were obtained by the procedure outlined previously and presented in Table II. The average values of the kinetic coefficients have been used to perform model simulations. These parameters were applicable to the data obtained from both the suspension and the fixed film bed reactor. Only the initial concentration of biomass was different. It was 4 times more in the fixed film reactor than in the free suspensions. This was reasonable since the support retained more biomass than a conventional reactor. The results of the simulation are for the suspension culture in Figure 2 and for the fixed film reactor in Figure 3. The agreement between observed and predicted concentrations is evident.

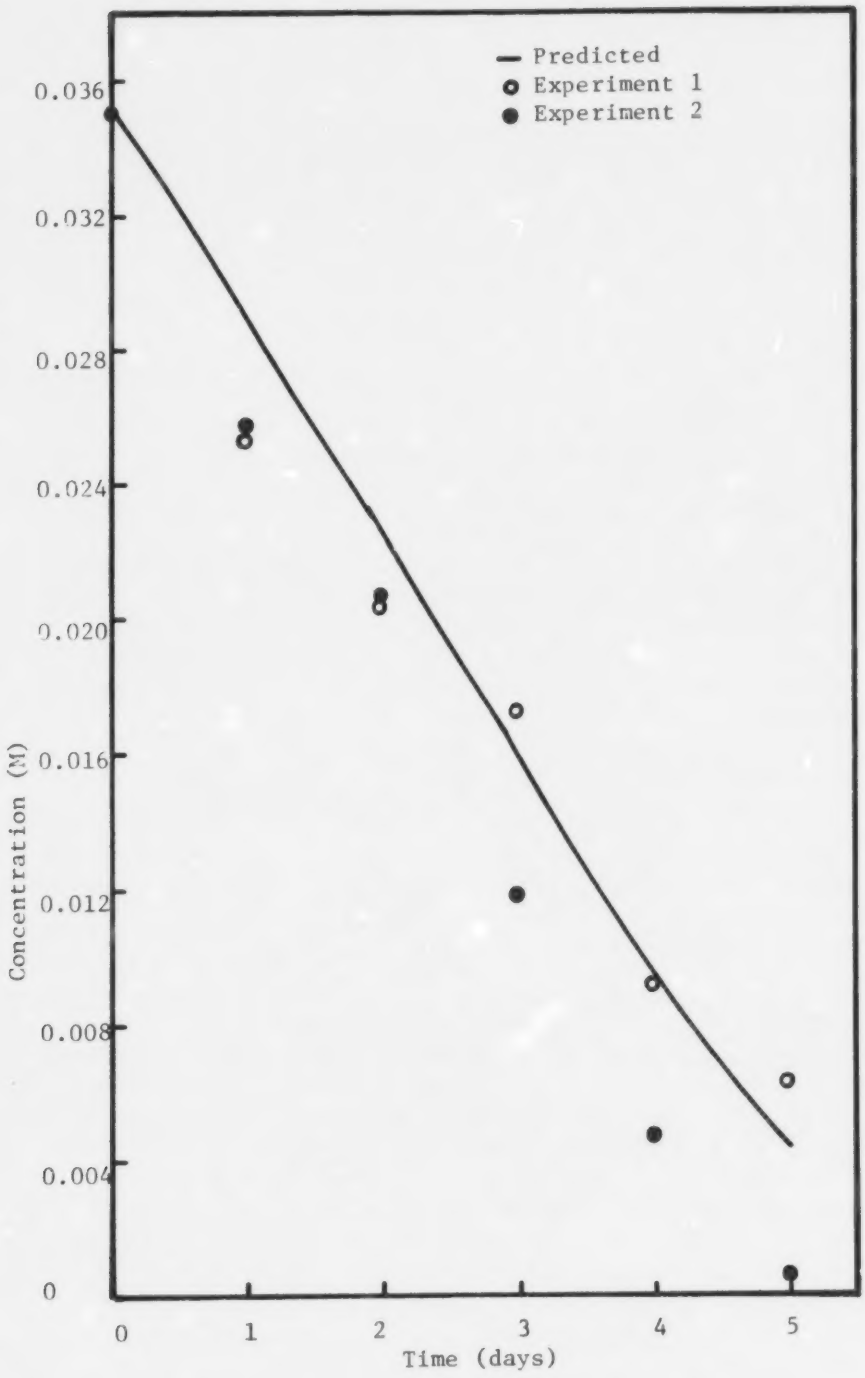


FIGURE 2. OBSERVED AND PREDICTED ACETATE UTILIZATION IN FREE SUSPENSIONS T = 39°C INITIAL pH = 7 UNCONTROLLED.

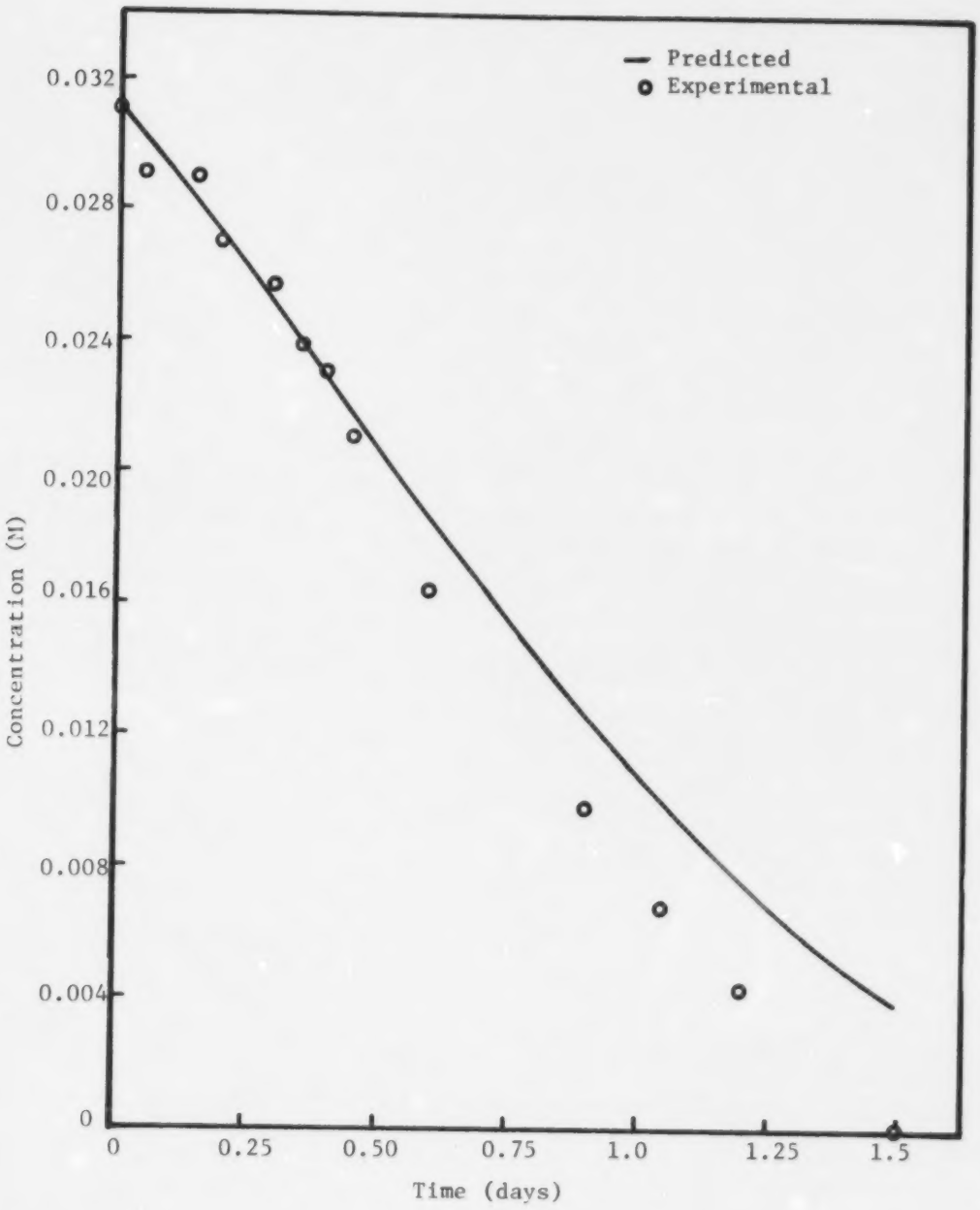


FIGURE 3. OBSERVED AND PREDICTED ACETATE UTILIZATION IN THE FIXED FILM BED REACTOR T = 39°C, pH = 7 CONTROLLED

TABLE II. KINETIC COEFFICIENTS USING FOR SIMULATION WITH THE TOTAL SUM OF SQUARES

Kinetic Coefficients	
μ_{max} = 0.17 day^{-1}	
K_{sa} = 8.46 x 10^{-3} M	
Xo/Yg = 0.0417 M	
k_p = 1.372 x 10^{-3} M·day^{-1}	
K_{sp} = 1.765 x 10^{-7} M	
k'_p = 0.5936	
Free Suspensions Experiment	**Total Residual Sum of Squares**
Acetate 1st Experiment	7.26 x 10^{-5}
Acetate 2nd Experiment	2.33 x 10^{-5}
Acetate + Propionate 1st Experiment	4.42 x 10^{-6}
Acetate + Propionate 2nd Experiment	1.46 x 10^{-6}
Fixed Film Reactor Experiment	
Acetate	5.25 x 10^{-5}
Acetate + Propionate	1.42 x 10^{-4}

The rate of acetate utilization expressed by equation 2 is in agreement with previous work (Ghosh and Klass, 1978 and Smith and Mah, 1978). Acetate does not appear to be inhibitory at high concentrations as indicated by Andrews and Graef (1971) and Hill and Barth (1977).

The model was applied to some previously published data using conventional reactor and fixed film bed reactor. The data reported by Summers and Bousfield (1976) using poultry waste in conventional reactor has been simulated as shown in Table III. The agreement is evident with the exception of 10 days retention.

TABLE III. OBSERVED AND PREDICTED ACETATE CONCENTRATIONS IN CONTINUOUS SUSPENDED CULTURE (DATA FROM SUMMERS AND BOUSFIELD, 1976)

T = 35°C Poultry Waste pH = assuming near neutrality

Retention Time (days)	Effluent Concentration (M as acetic acid)	Predicted Concentration (M as acetic acid)
10	4.57×10^{-3}	1.0×10^{-5}
7	1.54×10^{-2}	1.34×10^{-2}
5	1.91×10^{-2}	5.19×10^{-2}
3	4.95×10^{-2}	6.11×10^{-2}

TABLE IV. OBSERVED AND PREDICTED ACETATE CONCENTRATIONS IN FIXED FILM BED REACTOR (DATA FROM VAN DEN BERG AND KENNEDY, 1982)

pH = 7

Time (days)	Total Volatile Fatty Acids Conc. (M as acetic acid)	Predicted Concentration (M as acetic acid)
0.000	3.00×10^{-2}	3.00×10^{-2}
0.208	1.50×10^{-2}	1.81×10^{-2}
0.416	6.70×10^{-3}	8.19×10^{-3}
0.625	3.30×10^{-3}	2.30×10^{-3}

The data reported by Van Den Berg and Kennedy (1982) using pear-peeling in downflow fixed film bed reactor has been compared as shown in Table IV. the concentration of biomass was estimated from the division between the amount of acetate utilized per day in their experiments and the amount of acetate utilized per day measured in this experiment. The good agreement between observed data and predicted data is evident. From these results it could be concluded that the concentration of active biomass is several times greater than in a conventional reactor.

Acetate and Propionate Utilization

By examining the fate of propionate, in figures 4, 5 and 6 in the presence of acetate indicated that propionate utilization is inhibited by acetate. The application of model discrimination for the experiments in free suspensions indicated that competitive inhibition by acetate was the best model as evidenced by the lowest residual sum of squares (see Table V). Accordingly, the rate of propionate utilization in the presence of acetate can be expressed as follows:

$$\frac{dP}{dt} = \frac{-k_p P}{K_{sp} + P + A/k'_p} \qquad (3)$$

The different kinetic models for predicting acetate and propionate support, albeit indirectly, the presence of two distinct groups of microorganisms involved in the fate of acids in anaerobic digestion. Some assumptions have been done before estimating the parameters and simulating the experimental results. Propionate is non-methanogenic substrate. It can be assumed that the maximum production of acetate from propionate by the acetogenic bacteria will be 1.75 moles of acetate as shown below:

$$CH_2 CH_3 COO^- + 0.5\ HCO_3^- \rightarrow 1.75\ CH_3 COO^- + 0.25\ H^+ \qquad (4)$$

TABLE V. MODEL DISCRIMINATION FOR PROPIONATE IN PRESENCE OF ACETATE

Rate of Propionate Utilization (dP/dt)	Total Residual Sum of Squares 1st Experiment (Figure 4)	Total Residual Sum of Squares 2nd Experiment (Figure 5)
$\frac{-k_p P}{K_{sp} + P} e^{-k'_p A}$	5.04×10^{-5}	8.79×10^{-5}
$\frac{-k_p P}{K_{sp} + P}$	3.17×10^{-5}	4.33×10^{-5}
$\frac{-k_p P}{K_{sp} + P + A/k'_p}$	4.10×10^{-6}	1.14×10^{-6}

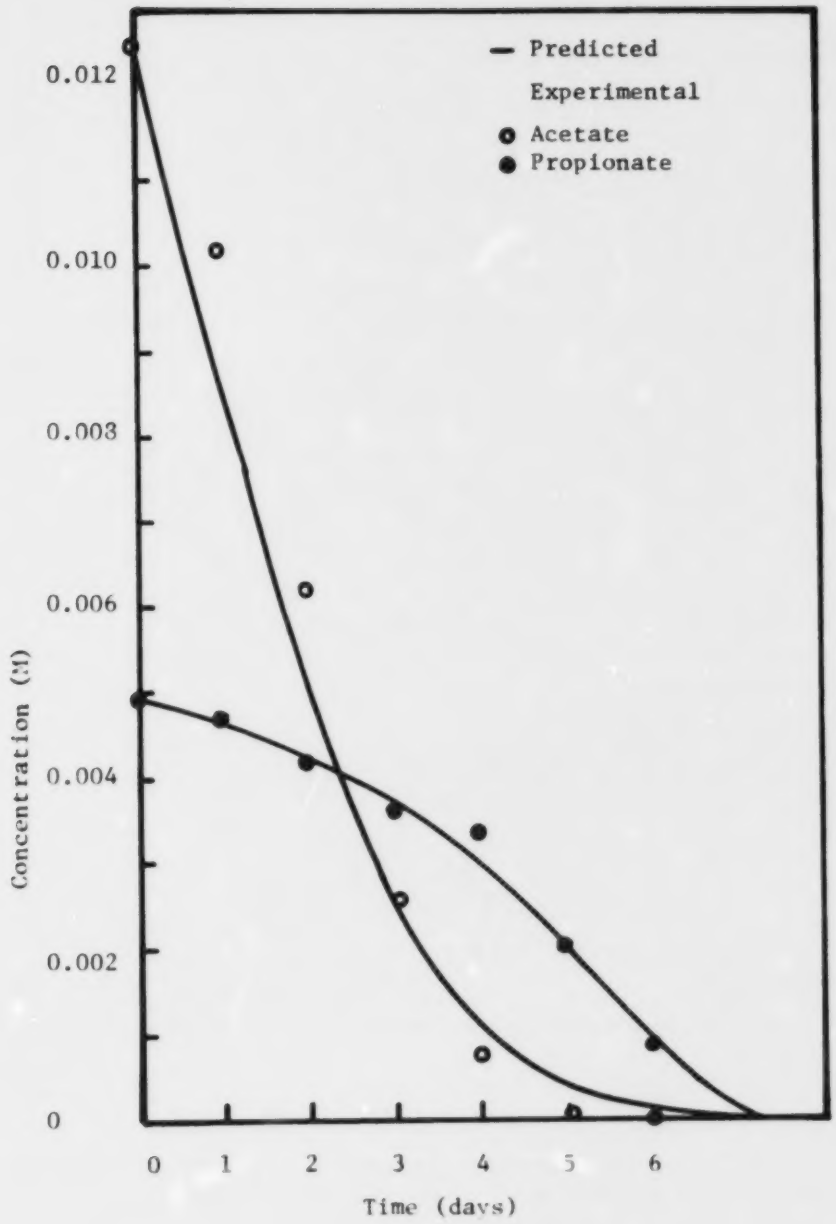


FIGURE 4. OBSERVED AND PREDICTED ACETATE AND PROPIONATE UTILIZATION (1ST EXPERIMENT) IN FREE SUSPENSION. T = 39°C INITIAL pH = 7 UNCONTROLLED

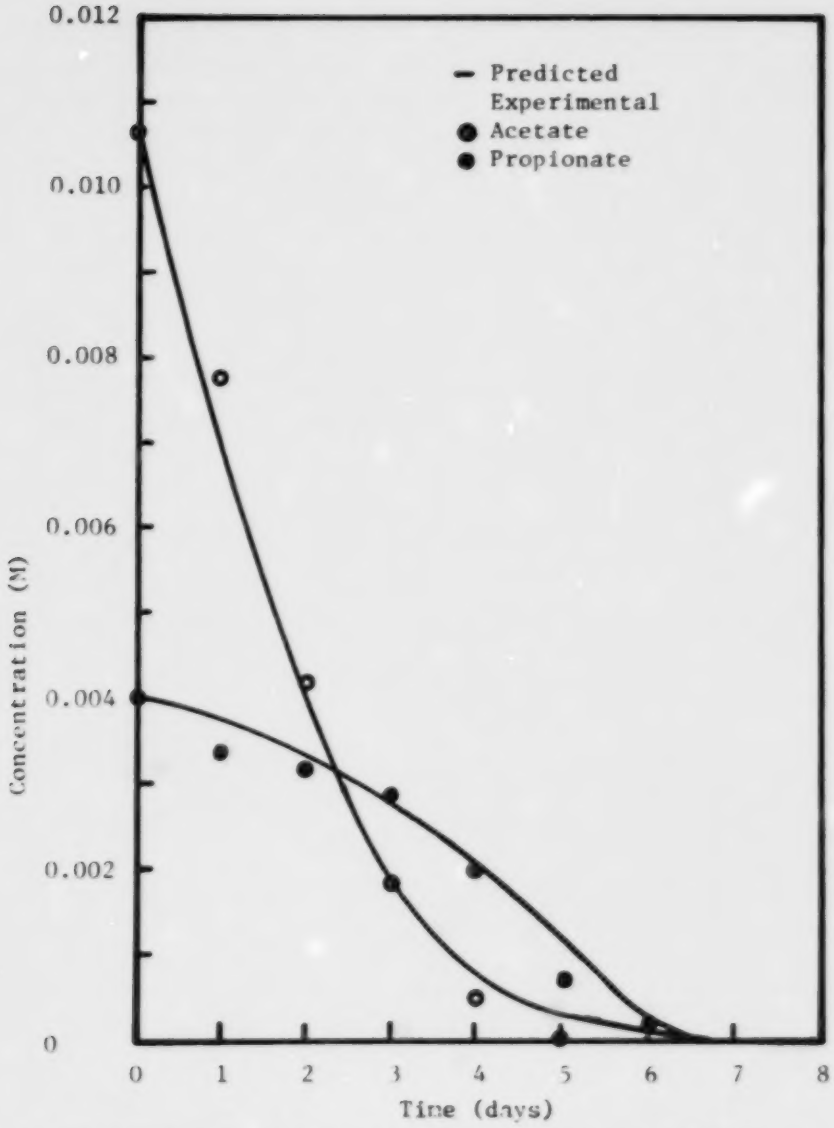


FIGURE 5. OBSERVED AND PREDICTED ACETATE AND PROPIONATE UTILIZATION (2ND EXPERIMENT) IN FREE SUSPENSION T = 39°C INITIAL pH = 7 UNCONTROLLED

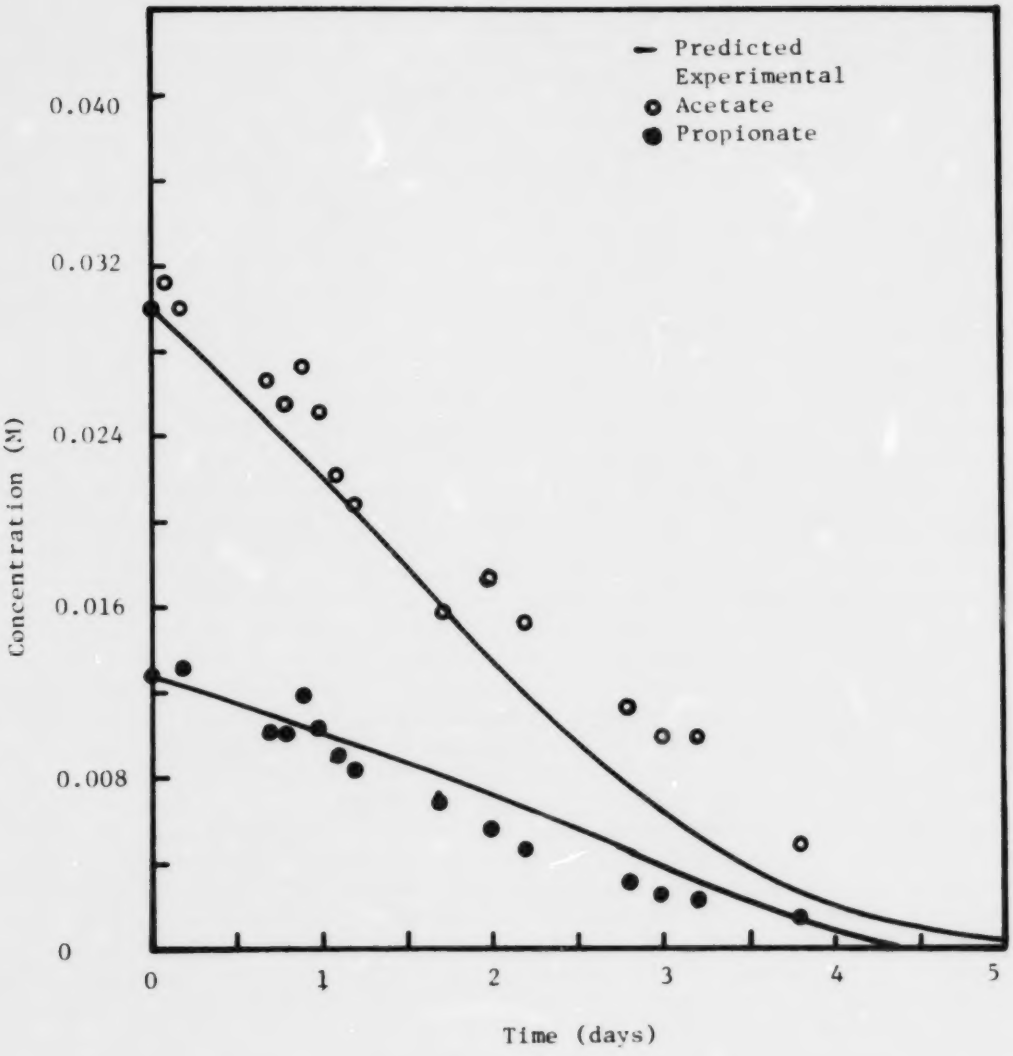


FIGURE 6. OBSERVED AND PREDICTED ACETATE AND PROPIONATE UTILIZATION IN FIXED FILM BED REACTOR. T = 39°C pH = 7 CONTROLLED.

Consequently, the differential equations used to estimate the parameters and to perform the simulation are as follows:

$$\frac{dA}{dt} = \frac{-\mu_{max} A\left[(A_o - A) + \frac{Xo}{Yg}\right]}{(K_{sa} + A)} + 1.75 \frac{k_p P}{K_{sp} + P + A/k'_p} \qquad (5)$$

$$\frac{dP}{dt} = \frac{-k_p P}{K_{sp} + P + A/k'_p} \qquad (6)$$

The parameters have been estimated as discussed in the previous section and are identified in Table II. The results of the simulation are for the suspension culture in Figure 4 and 5 and in Figure 6 for the fixed film reactor. The simulation, for the fixed film reactor, has been performed for an initial concentration of biomass 1.5 times greater than the initial one with the free suspensions and a specific rate of propionate 9.5 times greater than the value obtained with free suspension. The good agreement between observed and predicted concentrations is evident. In this study, the rate of propionate utilization was found to follow a competitive inhibition pattern. In this study the maximum specific growth rate couldn't be determined independently because the concentration of acetogenic bacteria did not change significantly during an experimental run. Furthermore, the acetogenic organisms are not easily differentiated from the methanogens. In this work, k_p the specific rate of propionate utilization is a combination of the specific growth rate (μ), the acetogenic biomass concentration (X_1) and the growth yeild of acetogenic bacteria (Y_{g1}). The model was applied to some previously published data. The data reported by Summers and Bousfield (1980) using swine waste in conventional reactor has been simulated in Table VI. The agreement is evident with the exception of acetate at 10 days retention. The data reported by Kennedy and Van Den Berg (1982) using piggery waste in fixed film bed reactors operated downflow has been compared with predictions, as shown in Table VII. The agreement between observed and predicted concentrations is good. The simulation has been performed for an initial concentration of biomass 5 times greater than in free suspensions and a specific rate of propionate 26.5 times than the value estimated from the free suspensions. It is interesting to note that the ratio between the specific rate of propionate and the initial concentration of biomass are 5.3 to 1 for our fixed film bed and 6.3 to 1 for Kennedy and Van Den Berg (1982) reactor's.

TABLE VI. OBSERVED AND PREDICTED ACETATE AND PROPIONATE CONCENTRATIONS IN CONTINUOUS SUSPENDED CULTURE (DATA FROM SUMMERS AND BOUSFIELD, 1980).

T = 35°C pH = near neutrality Swine Waste

Retention Time (days)	3 days Effluent Conc. (M)	3 days Predicted Conc. (M)	10 days Effluent Conc. (M)	10 days Predicted Conc. (M)
Acetate	1.42×10^{-2}	2.33×10^{-2}	1.28×10^{-2}	3×10^{-5}
Propionate	1.41×10^{-2}	1.15×10^{-2}	2.52×10^{-3}	5.18×10^{-3}

TABLE VII. OBSERVED AND PREDICTED ACETATE AND PROPIONATE CONCENTRATIONS IN FIXED FILM BED REACTOR (DATA FROM KENNEDY AND VAN DEN BERG, 1982)

T = 35°C Retention time = 2.6 days pH = 7.9 - 8.1 Substrate = Piggery Waste

Time (days)	Acetate Conc. (M)	Predicted Conc. (M)	Propionate Conc. (M)	Predicted Conc. (M)
0.000	2.92×10^{-2}	2.92×10^{-2}	6.76×10^{-3}	6.76×10^{-3}
0.083	2.67×10^{-2}	2.68×10^{-2}	6.43×10^{-3}	6.37×10^{-3}
0.250	2.08×10^{-2}	2.25×10^{-2}	5.41×10^{-3}	5.62×10^{-3}
0.417	1.67×10^{-2}	1.82×10^{-2}	4.73×10^{-3}	4.82×10^{-3}
0.583	1.25×10^{-2}	1.42×10^{-2}	4.05×10^{-3}	4.13×10^{-3}
0.750	1.00×10^{-2}	1.05×10^{-2}	2.97×10^{-3}	3.09×10^{-3}
0.833	8.33×10^{-3}	8.81×10^{-3}	2.70×10^{-3}	2.80×10^{-3}
0.917	8.00×10^{-3}	7.25×10^{-3}	2.57×10^{-3}	2.25×10^{-3}

Economical Evaluations

A preliminary economic analysis was performed for comparing a single stage conventional anaerobic digester with a two-stage system. To investigate the effect of improved C/N ratio on the economy, the addition of pulpmill sludge to the manure was also considered. Thus, three scenarios have been investigated as illustrated in Figure 7. These include conventional digestion in one stage, a two-stage system and two stages with addition of pulpmill sludge.

The major capital equipment cost (CEC) were estimated from the correlation of Hashimoto et al. (1980) or from Peters and Timmerhaus (1980) using the Marshall and Stevens indices to update the cost when necessary. All costs are converted to 1983 Canadian dollars. The fixed capital equipment (FCI) was found from the relationship: FCI = 2 x CEC to account for installation, labor, insulation, piping, electrical work, building and instrumentation. A factor of 2 was used to account for the low technology in a farm scenario. The return on investment (ROI) as outlined by Jelen (1970) was the prime parameter used for the economic analyses. To account for scale effects, the following standard approach was used.

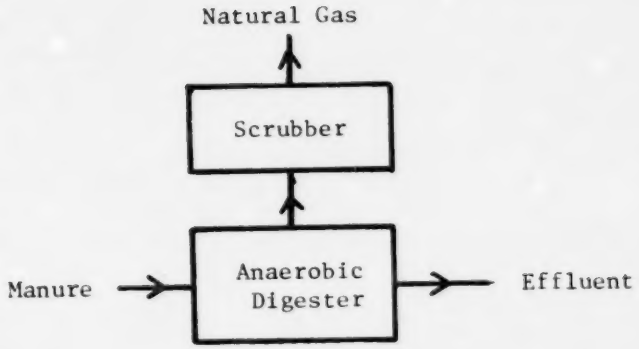
$$CEC_2 = CEC_1 \left[\frac{Capacity_2}{Capacity_1}\right]^{0.7}$$

A scale-up factor of 0.7 has been recommended by Hashimoto et al. (1980) for farm based anaerobic fermentor. With the exception of labour, all costs were scaled linearly with capacity. The labor costs were not varied with the capacities.

The biogas was cleaned in order to meet the quality standard of a regular natural gas supply company and no storage facility cost was assigned. Intermittent mixing of the digester contents was assumed, based on experimental evidence from our laboratory and from Haga et al. (1979) viz. half-hour period every 2 hours per 24 h-day. After digestion, the liquid withdrawn was assumed to meet environmental government standards and was conveniently discharged to a holding pond for use as a fertilizer. No transportation cost were assigned to pulpmill sludge or manure. Fertilizer value and solids fraction as compost were not credited. Details of the assumptions used in the economic evaluations are summarized in Table VIII.

The rate of acid production and gas production have been calculated theoretically from the composition of the swine manure. The solids composition of piggery waste, the concentration of acetate in the liquid fraction, and the assumptions made are identified in Table IX. The percentage of total solids was assumed to be 9.2% (ASAE D384, 1977). The lignin and the mineral fraction, of course were not convertable to biogas. The percentage of cellulose utilization was assumed to be 50% due to the lignin content. (Scharer and Moo-Young, 1979). This value is in agreement with the experimental value reported by Scharer et al. (1982). The rate of cellulose hydrolysis was assumed to be 1.2864 g/L•d (Scharer and Moo-Young, 1979). Tentatively, the digestible protein was assumed to be equivalent to casein; the lipid fraction was assumed to be equivalent to oleic acid (Metzler, 1977).

Case I



Case II and III

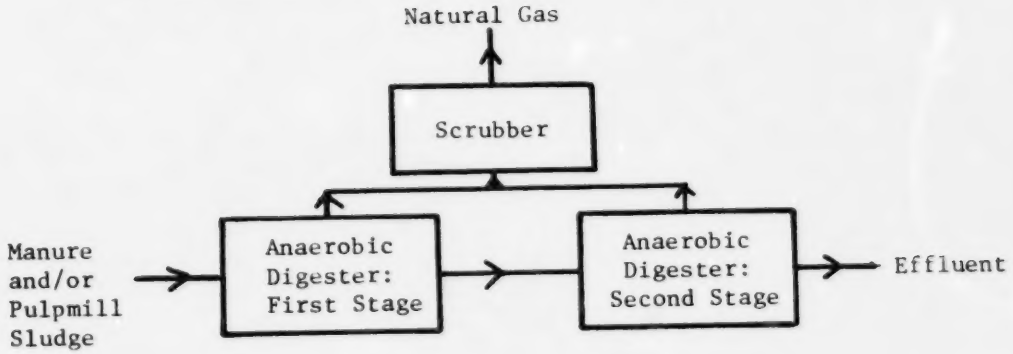


FIGURE 7. PROCESS OUTLINE FOR METHANE PRODUCTION

TABLE VIII. BASES FOR ECONOMIC EVALUATIONS

1. Fixed Capital Investment (FCI) = 2x Capital Equipment Cost
2. Maintenance = 1% FCC.
3. Depreciation Rate = 6.7% straight line.
4. Scale-up Factor: 0.7
5. ROI - Return on Investment Before Taxes = $\frac{\text{Operating Profit}}{\text{Total Capital Investment}}$
6. Direct Labour Cost = 7 days/week; 1 shift/day; 1 operator/shift; 8 h/shift; $7/h.
7. Continuous Operation 330 days per year.
8. Electricity = 4 cents/k Wh
9. Cost of Natural gas = $5.00/GJ
10. Raw material and utilities scale up linearly with plant capacity.

The stoichiometry of methane production in one stage has been proposed as follows:

Cellulose

$$C_5H_{10}O_5 + 0.32\ NH_3 + 0.04\ H_2O \rightarrow 0.32\ C_5H_7O_2N + 1.65\ CH_3COOH + 0.55\ CO_2 + 0.55\ CH_4$$

Protein

$$C_8H_{12}O_3N_2 + 4.58\ H_2O \rightarrow 0.14\ C_5H_7O_2N + 2.75\ CH_3COOH + 1.86\ NH_3 + 0.9\ CH_4 + 0.9\ CO_2$$

Lipid

$$C_{18}H_{34}O_2 + 0.4\ NH_3 + 14.8\ H_2O \rightarrow 0.4\ C_5H_7O_2N + 6\ CH_3COOH + 15\ H_2 + 2\ CH_4 + 2\ CO_2$$

Acetate

$$CH_3COOH + 0.032\ NH_3 \rightarrow 0.032\ C_5H_7O_2N + 0.92\ CH_4 + 0.92\ CO_2 + 0.096\ H_2O$$

TABLE IX. SOLIDS COMPOSITION IN SWINE WASTE (Holson, et al., 1974)

Ash	17.6%
Lignin	10.1%
Cellulose	22.9%
Hemicellulose	20.8%
Protein	20.9%
Lipid	7.7%

Concentration of acetate in the liquid fraction = 11.8 g/L (Georgacakis, et al., 1982)

Assumptions

% TS in swine manure = 9.2% (ASAE D384 1977)

Ash and lignin are not utilized

Rate of cellulose hydrolysis: 1.2864 g/L.day (Scharer and Moo-Young, 1979)

C/N ratio : 11 to 1

Protein equivalent to casein; casein formula = $C_8H_{12}O_3N_2$

Lipid equivalent to oleic acid; oleic acid formula = $C_{18}H_{34}O_2$

The percentage of cellulose utilization was assumed to be 50%; and protein and lipid: 35% as reported by Scharer et al. (1982). The acetate was assumed to be completely converted in methane and carbon dioxide. The gas production rate was found to be 1.21 $L/L_v \cdot d$ for a mesophilic reactor at 16 days retention time. The gas composition was assumed to be 50% CH_4 and 50% CO_2. The possible conversion of H_2 from the lipids to methane was not considered. However, conversion of H_2 from the lipids considered, the new gas composition would be 61.7% CH_4 and 38.2% CO_2. The gas production rate, using the correlation reported by Scharer et al. (1982) is predicted to be 1.20 $L/L_v \cdot d$ and they reported a gas compositon of 63% CH_4 and 37% CO_2 using swine manure under mesophilic conditions. Thus, the agreement between observed and predicted gas generation is excellent. The digestion conditions for the conventional reactor are tabulated in Table X.

The stoichiometry of methane production in the two-stage unit is as follows:

Cellulose

$$C_6H_{10}O_5 + 0.32\ NH_3 + 0.04\ H_2O \rightarrow 0.32\ C_5H_7O_2N + 2\ CH_3COOH + 0.2\ CO_2 + 0.2\ CH_4$$

Protein

$$C_8H_{12}O_3N_2 + 4.58\ H_2O \rightarrow 0.14\ C_5H_7O_2N + 1.86\ NH_3 + 3.325\ CH_3COOH + 0.325\ CO_2 + 0.325\ CH_4$$

Lipid

$$C_{18}H_{34}O_2 + 0.4\ NH_3 + 14.8\ H_2O \rightarrow 0.4\ C_5H_7O_2N + 15\ H_2 + 7.275\ CH_3COOH + 0.725\ CO_2 + 0.725\ CH_4$$

Acetate

$$CH_3COOH + 0.032\ NH_3 \rightarrow 0.032\ C_5H_7O_2N + 0.92\ CH_4 + 0.92\ CO_2 + 0.096\ H_2O$$

TABLE X. BASIC CONDITIONS AND PERFORMANCE DATA OF THE CONVENTIONAL REACTOR

Temperature	39°C
Retention time	16 days (continuous)
Gas Production Rate	1.21 $L/L_V \cdot d$
Gas Composition (experimentally)	63% CH_4, 37% CO_2

In this case more acetate has been produced than the conventional reactor. This has been due to the more favorable conditions for acetogenic bacteria in the first stage. In the first stage, the percentage of cellulose utilization was assumed to be 50% while protein and lipid use 100%.

The retention time for the acidogenic stage has been determined by dividing the amount of cellulose to be utilized over the rate of cellulose hydrolysis. Protein and fat were assumed to have a higher rate of hydrolysis than cellulose. The rate of acid productions has been calculated to be 0.11 M acetate/day. The amount of acetate utilized in the first stage was assumed to be negligible. The gas production was found to be 0.56 $L/L_V \cdot d$ (50% CH_4 and 50% CO_2). The second stage was assumed to be a downflow fixed film bed reactor. As before, all acetate in the liquid fraction was assumed to be converted to methane and carbon dioxide. The retention time has been determined by dividing the amount of acetate to be utilized over the rate of

acetate utilization. The gas production was found to be 4.37 $L/L_v \cdot d$ By assuming an efficiency of 80% for the bioreactor, the gas production would be 3.5 $L/L_v \cdot d$ which is in agreement with the value reported by Kennedy and Van Den Berg (1982). The gas composition was assumed to be 72% CH_4 and 28% CO_2 (Kennedy and Van Den Berg, 1982). The digestion conditions for the two stages are listed in Table XI.

TABLE XI. BASIC CONDITIONS AND PERFORMANCE DATA OF THE TWO-STAGE SYSTEM

I. First Stage: Acid Production	
Temperature	39°C
Retention time	8.3 days
Gas production rate	0.56 $L/L_v \cdot d$
Gas composition (theoretically)	50% CH_4, 50% CO_2
II. Second Stage: Acid Utilization	
Temperature	39°C
Retention time	5.5 days
Gas production rate	4.37 $L/L_v \cdot d$
Gas composition (experimentally)	72% CH_4, 27% CO_2

Research in our laboratory and published results by Hills (1979) showed that the optimum C/N ratio for anaerobic digestion is 24/1. Since the C/N ratio in swine manure is only 11/1 the addition of a relatively cheap digestible, high carbon source was considered. Pulpmill sludge would meet these requirement. Approximately 0.9 tonnes (dry weight) of cellulose from pulpmill sludge per tonne (dry weight) of manure was necessary to bring the C/N ratio to 24/1. The calculation procedure has been described previously. The digestion conditions for the two stages with pulpmill sludge are tabulated in Table XII.

TABLE XII. BASIC CONDITIONS AND PERFORMANCE DATA OF THE TWO STAGE SYSTEM WITH ADDITION OF PULPMILL SLUDGE

I. First Stage: Acid Production	
Temperature	39°C
Retention time	12.2 days
Gas production rate	0.59 $L/L_v \cdot d$
Gas composition (theoretically)	50% CH_4, 50% CO_2
II. Second Stage: Acid Utilization	
Temperature	39°C
Retention time	5.8 days
Gas production rate	5.93 $L/L_v \cdot d$
Gas composition (experimentally)	72% CH_4, 27% CO_2

Case I — Conventional Reactor

Figure 8 shows the results for conventional reactor. In this case, a positive return on investment at the price of natural gas ($5/GJ) can be obtained at 120.5 tonnes/day, which is equivalent to manure obtained from approximately 147,175 swines or 16,350 dairy cows. From these results it may be concluded that cooperative programs should be considered in order to make the process economically viable. A comparison of the distribution of the major cost components in Table XIII shows that utilities (39.4% of the total cost) and equipment depreciation (58.4%) contribute the most to the total production cost.

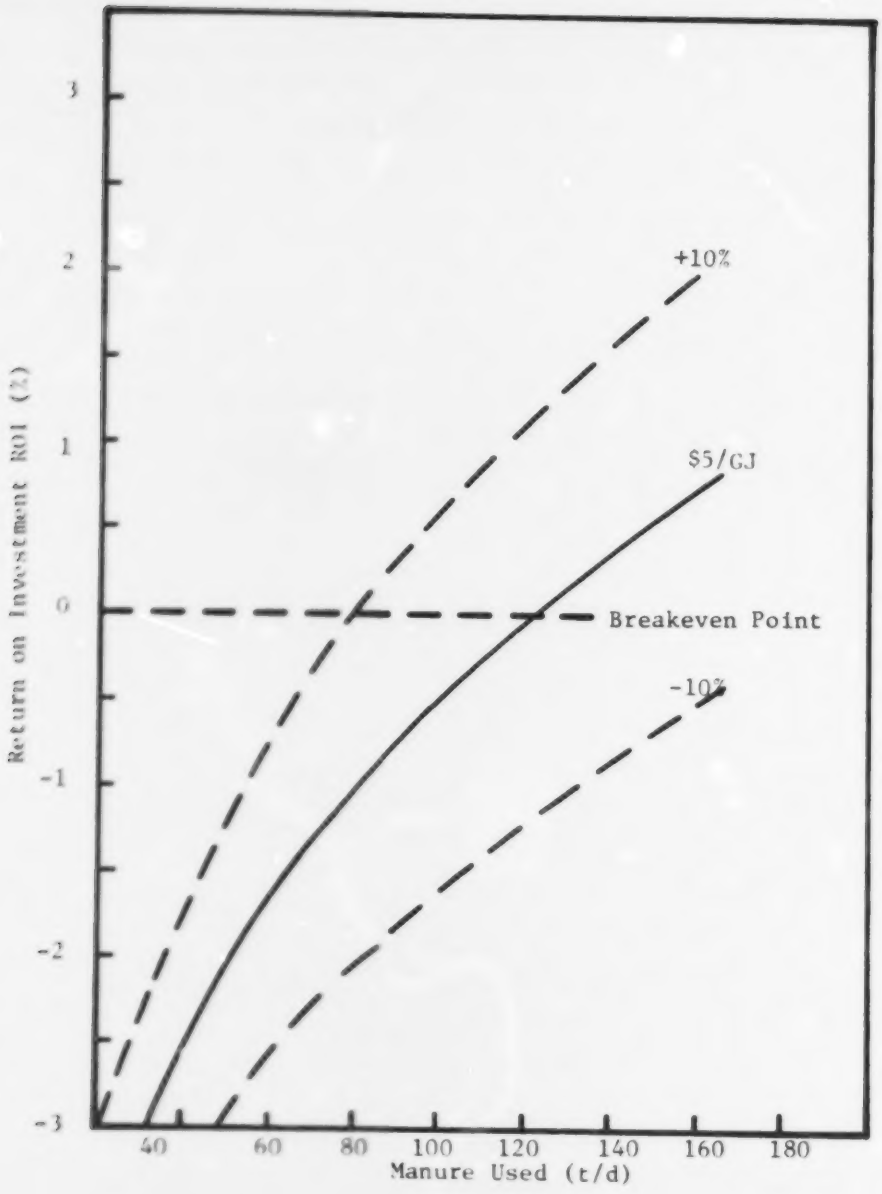


FIGURE 8. EFFECT OF SCALE ON ECONOMICS OF METHANE PRODUCTION WITH CONVENTIONAL REACTOR

TABLE XIII. COST DISTRIBUTION FOR CONVENTIONAL REACTOR AT BREAKEVEN POINT

Component	Cost ($)	% of Total Cost
Raw Materials	0	[0]
Operating Cost	383952.07	[41.6]
Utilities	364352.07	(39.4)
Labour	19600.00	(2.2)
Equipment Depreciation	539749.87	[58.4]
TOTAL	923701.94	(100.0)

Case II Two-Stage Anaerobic Digestion

Figure 9 shows the effect of a two-stage anaerobic digester on the ROI. The process economics is improved by increasing the methane gas production. It is seen that approximately 21.65 tonnes manure/day (equivalent to 26450 swines or 2940 cows) is necessary to break even at the price of natural gas ($5/GJ). Thus, the two-stage anaerobic systems appears to be economically more attractive. The two major reasons are an increase in the gas production and methane content and a lower initial investment cost than conventional reactor. The distribution of cost in Table XIV shows that utilities (37.2% of the total cost) and equipment depreciation (55.2%) contribute the most to the total production cost.

TABLE XIV. COST DISTRIBUTION FOR THE TWO-STAGE SYSTEM AT BREAKEVEN POINT

Component	Cost	% Total Cost
Raw Material	0	[0]
Operating Cost	115468.72	[44.8]
Utilities	95868.72	(37.2)
Labour	19600.00	(7.6)
Equipment Depreciation	142143.47	[55.2]
TOTAL	257612.19	(100.0)

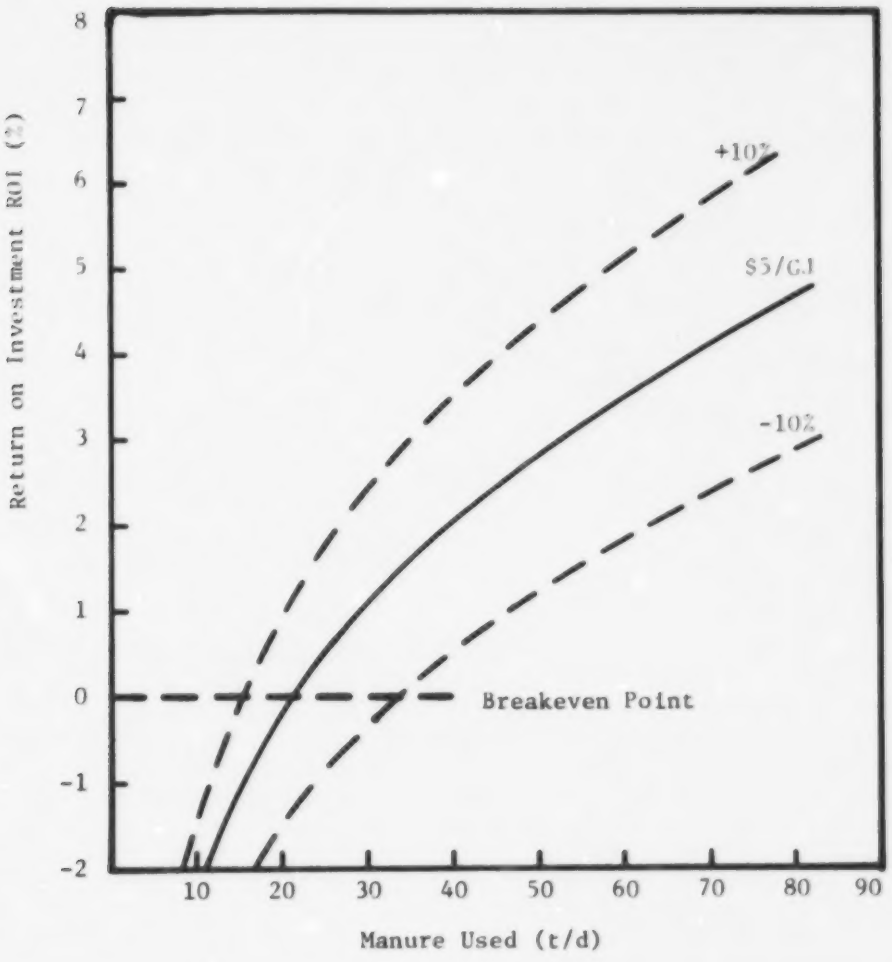


FIGURE 9. EFFECT OF SCALE ON ECONOMICS OF METHANE PRODUCTION WITH TWO-STAGE SYSTEM

Case III. Addition of Pulpmill Sludge to Two-Stage Anaerobic Fermentor

Figure 10 shows the effect of pulpmill sludge addition. At the price of natural gas ($5/GJ) a positive return on investment can be obtained at 14.1 tonnes per day (equivalent to 17300 swines or 1920 cows). The use of pulpmill sludge at zero cost may be not feasible in practice since it is unlikely that farms (swine or dairy cattle) may be located near pulp and paper mills. The trucking was not considered in this analysis. The distribution cost in Table XV shows that utilities (36.2% of the total cost and equipment depreciation (55.7%) contribute the most to the total production cost.

TABLE XV. COST DISTRIBUTION FOR THE TWO-STAGE SYSTEM WITH ADDITION OF PULPMILL SLUDGE AT BREAKEVEN POINT

Component	Cost	% of Total Cost
Raw Material	0	[0]
Operating Cost	108019.09	[44.3]
Utilities	88419.09	(36.2)
Labour	19600.00	(8.1)
Equipment Depreciation	135927.05	[55.7]
TOTAL	243946.14	(100.0)

Conclusion

In this report it has been shown that Monod kinetics can be used to predict the rate of acetate utilization and that acetate was not self inhibitory. Experimental results also showed that propionate utilization was inhibited by acetate and follow a competitive inhibition pattern.

The agreement between observations and predictions support the hypothesis that acetogenic bacteria produce 1.75 moles of acetate from 1 mole of propionate. Different kinetic models for acetate and for propionate support the current understanding that acetogenic bacteria, which belongs to a metabolic group different from methanogenic bacteria, were involved during the fermentation of propionate.

This dynamic model was applicable to previously published data on fixed film bed reactors. Good correlation between observed and predicted data was found.

From the preliminary economical evaluations it was concluded that the two-stage system is economically more attractive than a conventional reactor. Using a carbon to nitrogen ratio of 24:1 the two-stage system could be economically feasible at medium farm size.

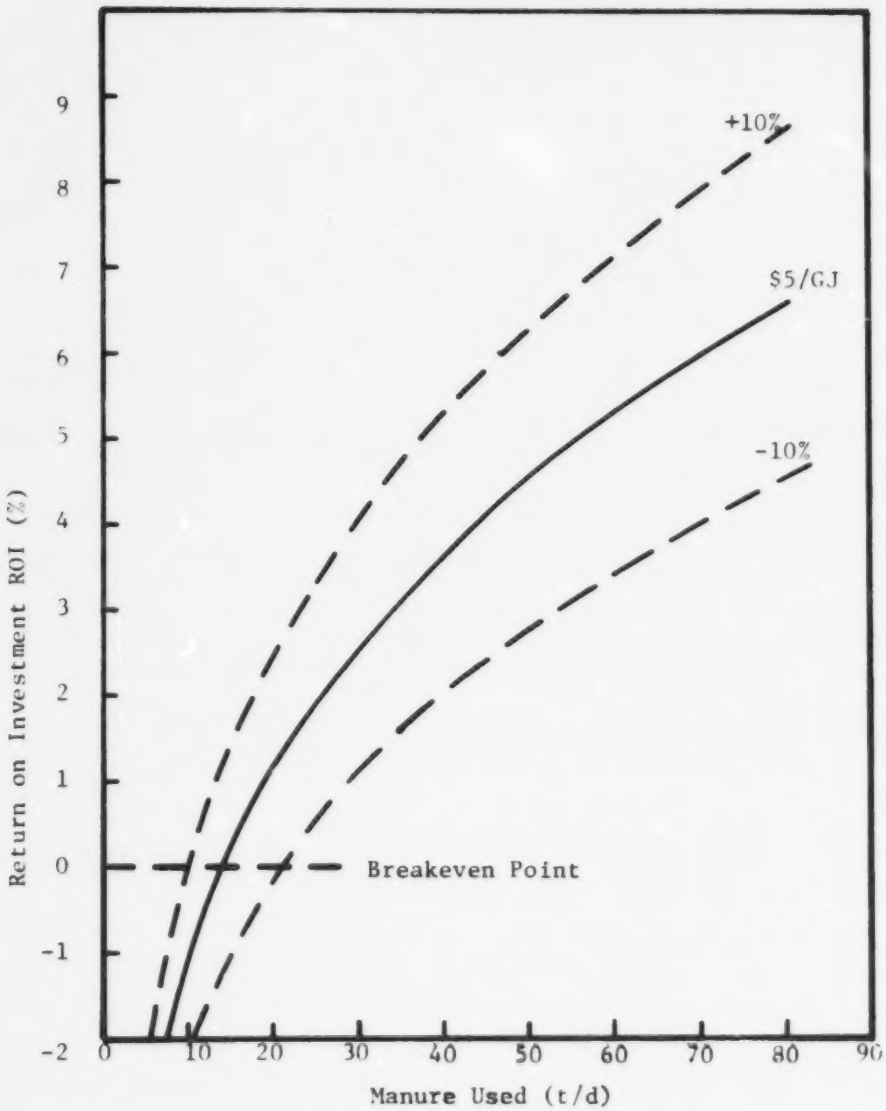


FIGURE 10. EFFECT OF SCALE ON ECONOMICS OF METHANE PRODUCTION WITH TWO-STAGE SYSTEM WITH ADDITION OF PULPMILL SLUDGE.

Notations

A = concentration of acetate (M)

A_0 = initial concentration of acetate (M)

$\frac{dA}{dt}$ = rate of acetate utilization ($M \cdot day^{-1}$)

$\frac{dP}{dt}$ = rate of propionate utilization ($M \cdot day^{-1}$)

K_{sa} = Monod constant of acetate (M)

k_p = specific rate of propionate ($M \cdot day^{-1}$)

k'_p = inhibition constant for propionate

K_{sp} = saturation constant of propionate (M)

P = concentration of propionate (M)

X_0 = initial concentration of methane formers (M)

Y_g = growth yield of methane formers (mg/mg)

μ_{max} = maximum specific growth rate of the methane formers (day^{-1})

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RÉHABILITATION DE L'USINE D'ÉPURATION DE VAUDREUIL (Québec)

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USE OF AQUATIC MACROPHYTES FOR SLUDGE TREATMENT

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ABSTRACT

Emergent aquatic macrophytes are water tolerant vascular plants that can be used to greatly improve and expand the performance of conventional sludge drying beds. Reed beds are modified sand beds planted with reeds (Phragmites communis). The function of reeds is to provide components of the aquatic environment that improve the dewatering capability and/or reliability of conventional sludge drying beds without the use of chemicals or any subsequent energy intensive handling or processing techniques. Aerobic stabilized sludge is usually pumped directly from the aeration basin(s) to the reed beds. The deposited sludge initially undergoes air drying, i.e. moisture is lost to the atmosphere by evapotranspiration and to the bed by physical filtration where particulates are filtered mechanically as water passes through the sand filter, the root masses and decomposing dead organic matter.

In the southern part of Germany about 20 dryings of 8 - 12 cm of wet digested sludge are feasible per year on open beds. If enclosures protect the site from rain the number of dryings per year could be increased by 30 - 70 %.

Sludge is not removed from the reed beds after it has drained and dried sufficiently but remains there until secondary roots and shoots are able to assist in decomposing the sludge residue by means of bacterial and plant metabolism in the rhizosphere of Phragmites communis. The end product is a stabile humus-like material. The bed turnover time is at least 5 years. The oldest field scale plant at the WWT facilities in Ustersbach, Germany, employing 300 m^2 of reed beds has already been operating intermittently, i.e. 8 - 9 months per year, for 8 years without removing the humus build-up. Experience so far has shown that the sludge brought to the reed beds does not require any prethickening and/or chemical treatment. The advantages of reed beds over conventional facilities lie in the long bed turnover time and in the by-products (biomass and low grade compost) produced.

INTRODUCTION

The oldest and most widely used method of sludge dewatering is air drying of sludge on drying beds. According to Burd (1) over 6000 wastewater treatment facilities use this method in the United States. Although they are popular in small- and medium-sized communities, drying beds are also used by 38 percent of the cities serving populations of

over 100,000. Burd also mentions that it is the most common dewatering technique employed in Europe. However in large municipalities, the initial cost, the cost of frequently removing the sludge and replacing sand, unfavourable climatic conditions, and the large area requirements may preclude the use of conventional sludge-drying beds.

It will be shown that the application of fundamental principles from aquatic treatment technology is a significant step in improving the performance of drying beds. The use of emergent aquatic macrophytes for the treatment and disposal of digested or stabilized sludge is a relatively recent development. It dates back to the mid 1960's when Dr. K. Seidel began to investigate the applicability of aquatic macrophytes to dewater the chemical-precipitation sludge from a nuclear research station after pilot plant studies had shown that conventional dewatering processes were either uneconomical or not very effective. The results of this investigation were far better than those for conventional drying beds (2). On the basis of this experience several field scale pilot plants, termed herein "reed beds", were built in Austria and southern Germany to carry out additional studies with activated sludge.

The objectives of this paper are to review what use has been made of aquatic macrophytes to treat excess activated sludge. To do this, the material to be presented is organized into three sections: 1) an overview of the type of pilot plants installed, 2) the physical, chemical, and biological transformations occurring in reed beds, and 3) summarize information on the performance of existing field scale facilities.

SLUDGE TREATMENT IN FIELD SCALE REED BEDS

Reed beds are conventional sludge drying beds planted with reeds (Phragmites communis) and modified in their design to be able to maintain soggy filter conditions for substantial periods of time.

A typical cross section of a reed bed is shown in Figure 1. The treatment area is partitioned into individual beds, approximately 6m wide by 20 to 30 m long, or a convenient size so that the sludge can spread uniformly and that soggy conditions can be maintained throughout the bed.

Reed beds may be roughly categorized as rectangular beds with side walls, layers of sand and then gravel with an under drainage system to collect and carry away the filtrate. They are lined with impervious membranes and planted with the reed, Phragmites communis. They should be built with a provision for mechanical removal, and may be built with or without either a roof or a greenhouse type covering. The drainage system can either be underdrain piping or perforated tiles.

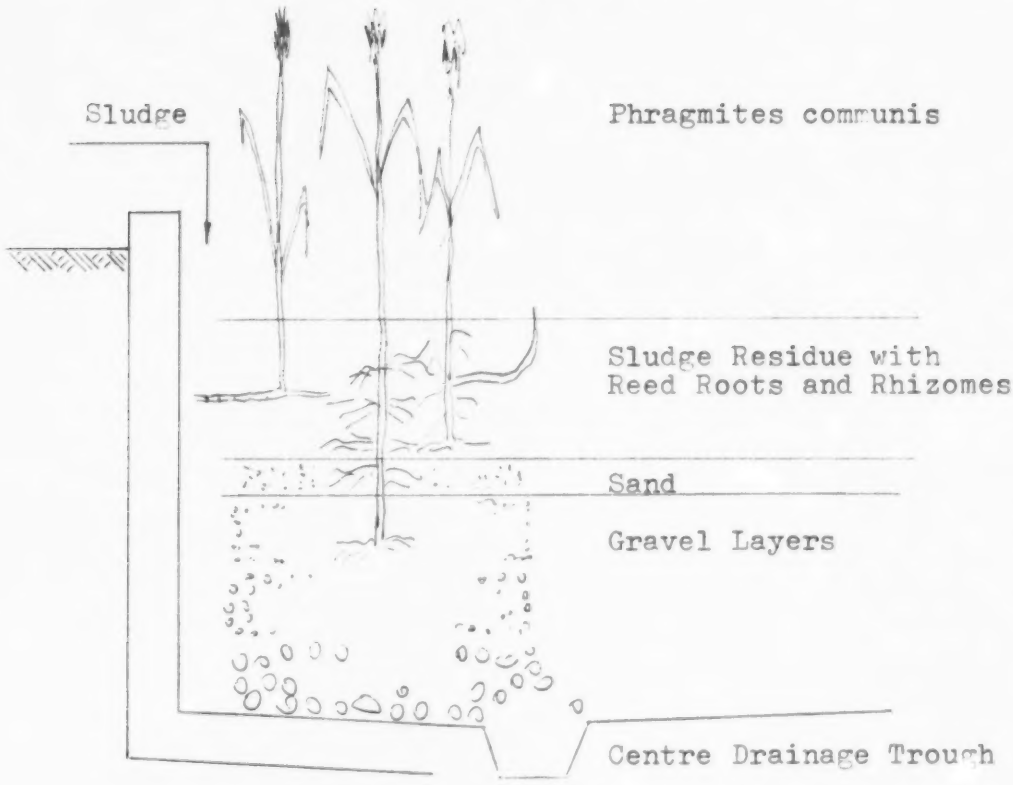


FIGURE 1. CROSS SECTION OF A TYPICAL REED BED

The bed drainage media normally consists approximately as follows:

1. The top layer is 5 cm of sand, usually with an effective size of 0.3 to 3 mm.

2. Three intermediate gravel layers, each 10 to 15 cm deep, having from the top to bottom an effective size of 4 to 8 mm, 16 to 32 mm, and 32 to 64 mm respectively.

3. Underdrain piping or perforated tiles should be of a material that does not fail when front end loaders are driven across the bed for solids residue removal.

TRANSFORMATIONS OCCURRING IN REED BEDS

The operation of reed beds is fairly simple. Stabilized sludge is conveyed to one or more modified drying beds planted with one species of aquatic macrophytes, namely Phragmites communis. This reed, itself, brings about very little actual treatment of the sludge. Its function is to provide components of the aquatic environment that improve the sludge

treatment capability and/or reliability of that environment. These functions are presented in Table 1.

TABLE 1. FUNCTION OF PHRAGMITES COMMUNIS IN REED BEDS

Plant Parts	Function
Roots and/or stems in the sludge column	1. Surfaces on which bacteria grow. 2. Media for filtration and adsorption of solids, nutrients and water.
Stems and/or leaves at or above the sludge surface	1. Attenuate sunlight and thus can prevent growth of suspended algae. 2. Evaporation of large amounts of water.

There are also other important reasons why of more than 200.000 known species of higher plants a reed was chosen as the most promising vegetation to assist in the dewatering and conversion of sludge. Studies (3,4) have shown that

1) Phragmites communis may be the most widespread of all flowering plants. It grows coast to coast from central Canada to southern South America, across Eurasia, in most parts of Africa and in Australia.

2) Phragmites communis is incredibly adaptible, growing up in mountains of Bolivia and Central Asia (5000 m) as well as in the hot and humid deltas of the Tigris and Euphrates.

3) Phragmites communis is a tough plant that can grow in polluted waters or thrive in sludges that would eliminate most other vegetation.

4) Phragmites communis is anchored in the soil by both roots and rhizomes, a network of underground stems from which new plant sprout.

5) Phragmites communis is tolerant to low oxygen levels and to anaerobic reactions in the rhizosphere, and even to waterlogged conditions provided the plant is not regularly submerged.

Reed beds are an aquatic system where sludge is treated principally by means of bacterial and plant metabolism and physical filtration. When stabilized sludge is delivered to reed beds the primary removal and separation mechanism for the solids and contaminants in the sludge is filtration where particulates are filtered mechanically as water passes through the bed drainage media, the roots and dense mat of rhizomes. Secondary removal mechanisms are sedimentation, adsorption and precipitation (5). Dewatering initially proceeds by drainage and then continues by evapotranspiration. This is significant because plant transpiration will continue the drying process when evaporation would have reached the constant rate phase of sludge drying.

Sludge is not only transformed physically and mechanically but also biologically. In the dense mat of rhizomes are attached micro organisms that together with root excretions will metabolize colloidal solids, soluble organics and some nutrients. To a lesser degree plant absorption during the growth phase will take up significant quantities of contaminants.

In addition to the removal of the constituents in the sludge, processes leading to the eventual decomposition of the organic matter deposited in the rhizosphere are of fundamental importance in the operation and management of reed beds. Two basic processes must be considered. They are mineralization and humification (6,7).

Mineralization occurs as a result of the metabolism of microorganisms. Metabolism refers to both biosynthesis, in which organic matter is assimilated into cell tissue, and to catabolism, in which organic matter is converted to simple compounds to obtain energy for cell synthesis and maintenance. Dead organic matter not used for cell production is released in the form of minerals or simple organic compounds.

Humification is the process by which organic compounds are transformed into a material called humus. The process of humification involves a long series of biochemical transformations in which a variety of organic compounds are slowly converted into complex organic heteropoly-condensates with bonds of different strengths. Bonding with mineral constituents in the environment such as metal ions in solution in the sludge affects both the formation and stability of humic substances.

The development of a humus-like material in the rhizosphere of reeds is especially important in the reduction and stabilization of the filtered residue.

In reed beds sludge treatment occurs in a comparatively slow rate in an essentially unmanaged natural environment. The consequence of this is that reedbeds are subject to lesser operational control and to more environmental influence. The lack of chemicals or energy necessary to dewater and convert the sludge should be a major stimulus to use reed beds.

PERFORMANCE OF EXISTING FIELD SCALE FACILITIES

To date the following field scale reed bed installations have been built:

1) Ustersbach, Germany, operating since 1974, 300 m^2 of reeds, aerobic stabilized sludge

2) Dornstadt , Germany, operating since 1981, 240 m^2 of reeds, raw, aerobic and anaerobic stabilized sludge

3) Neumarkt , Austria, operating since 1982, 280 m^2 of reeds, aerobic stabilized sludge

The Ustersbach reed beds are the oldest of the field scale installations. A smaller pilot plant of approximately 50 m^2 of reeds was used between 1965 and 1972 to dewater chemical-precipitation sludge (8,9).

Investigations carried out in 1979 and recommenced in 1982 at the Ustersbach wastewater treatment facilities will form the basis of this chapter and will be complemented with data from the other installations.

Description of the Ustersbach WWT Plant

The wastewater treatment plant in Ustersbach is an extended aeration process. Separate aerobic digestion facilities were shut down in 1971, so that excess sludge is removed directly from the return sludge line and sent either to a thickener or to the reed beds. The reed beds are operated as a pilot plant and receive approximately 15 % of all excess sludge generated. The remaining 85 % of the sludge is thickened and then discharged to neighbouring fields.

The design and operating characteristics of the wastewater treatment plant in Ustersbach are shown in Table 2. The source of the wastewater is the town of Usterbach and a nearby brewery. Because of the high organic waste load from the brewery the treatment system cannot be considered as a typical domestic sewage treatment plant. Most of the brewery organic waste is soluble so that the excess sludge production rate is less than that expected for domestic sewage.

Description of the Ustersbach Reed Beds

Three of the conventional paved rectangular drying beds with a centre drainage trough and concrete side walls were modified to meet the design criteria mentioned previously and then planted with Phragmites communis. Each bed has a separate sludge distribution pipeline that is suspended longitudinally over its center. The sludge discharges out of six outlets onto the bed surface.

A fourth unplanted bed, designed in a manner similar to the reed beds, was used as a control in order to identify the influence that reeds have on sludge treatment. Additional data is presented in Table 3.

TABLE 2. SUMMARY OF DESIGN CHARACTERISTICS OF WWT PLANT USTERSBACH

Item	Amount
Plant design data	
- max brewery flow rate	1000 m^3/d
- max community flow rate	378 m^3/d
- max brewery BOD load	1220 kg/d
- max community BOD load	180 kg/d
Operating data	
- max brewery flow rate	658 m^3/d
- avg brewery flow rate	334 m^3/d
- max community flow rate	140 m^3/d
- avg community flow rate	125 m^3/d
- max brewery BOD load	360 kg/d
- max community BOD load	60 kg/d
Aeration basin	
- type	complete mix
- flow	459 m^3/d
- volume	1160 m^3
- detention	2.5 days
- max BOD load	0.36 $kg/m^3/d$
- max F/M	0.10 kg/kg/d
Secondary clarifier	
- surface area	210 m^2
- volume	315 m^3
Sludge drying beds	
- number	12
- area/bed	100 m^2
- bed dimensions	5 m x 20 m

TABLE 3. SUMMARY OF DESIGN OF THE USTERSBACH REED BEDS

Item	Bed 1	Bed 2	Bed 3	Bed 4
Bed area, m^2	100	100	100	100
Bed dimensions w x l, m	5 x 20	5 x 20	5 x 20	5 x 20
Date planted	1974	1974	1976	----
Reed type	80 % import 20 % local	100 % local	100 % local	----

Sludge Loading Experiments

After operating for five years the first loading experiments were carried out to determine loading rates that would ensure a bed life time of at least 4 to 5 years.

The test results are summarized in Table 4 and show that the reed beds can receive higher sludge quantities than conventional drying beds. Under similar loading rates the volume of the sludge residue in conventional beds was almost 290 % higher than in reed beds. The high volume reduction increased when we compared the total sludge volume deposited since the beds were planted to the residual volume remaining to date. In other words between 1974 and 1982 the volume reduction remains at 98 % for bed 1, and increases to 98.3 % and 97.6 % for beds 2 and 3 respectively.

The TS of the sludge residue peaked at 46 % in the reed beds when for several weeks there was almost no precipitation.

The reed beds have never been loaded continuously. The first major interruption occurred between March 1975 and June 1976, i.e. nine months after the beds had been planted. During the first nine months the beds were loaded at a rate of 25 to 28 cm every 10 to 14 days. The sludge had an average TS of 9.2 g/l. With a volume reduction of only 93 % the beds would have had to be cleared at least every six months. The poor performance after plant start-up is attributed to the following:

1) The beds were loaded at a time when the reeds had not reached maturity. A dense mat of rhizomes was absent.

2) The beds were loaded at least 60 % above the loading rates that we consider feasible.

TABLE 4. SUMMARY OF SLUDGE LOADING EXPERIMENTS

Item	Bed 1	Bed 2	Bed 3	Bed 4
Loading Period	Feb to Aug	April to Aug	April to Aug	April to Aug
Number of Loads	16	11	11	11
Quantity per Load, m^3	10.3	13.7	17.2	10.3
Total Sludge Quantity Delivered, mm	1650	1510	1890	1130
Residual Sludge Height Increase(Avg) mm	35	55	60	65
Volume Reduction, %	98	96.4	97	94.2
Average TS of the Sludge, %	1	1	1	1
Average TS of the Sludge Residue, %	24	34	29	20
TS Loading Rate, kg/sq.m/a	30.7	42.2	52.7	31.5

Remarks:

1. The precipitation amounting to approximately 23 % of the total hydraulic load was not included in the volume reduction calculations.
2. The TS of the sludge residue is influenced by the weather It fluctuated between 12 % and 46 %.

3) Loading occurred mainly during the winter season when sludge dewatering is affected by low temperatures.

For these reasons the reed beds were taken out of operation until June 1976 when the sludge residue height had fallen from 35 cm to 12 cm, i.e. when a settlement of 66 % had been reached.

The second major interruption occurred between April 1978 and January 1979 when the extended aeration process was shut down frequently in order to carry out repairs.

Reed Bed Runoff

The four beds were loaded at approximately 10 day intervals with excess sludge drawn off from the return sludge line. The runoff of the filtrate was measured for each bed. An analysis of the four filtrate hydrographs could not conclusively identify a higher runoff rate when there is present a dense mat of rhizomes. The only observation that we can make so far is that after operating for eight years the bed drainage media as well as the large volume of sludge residue have remained unclogged. The roots and rhizomes are a living filter that prevents clogging. The drainage process is fairly rapid so that within 2 to 3 days approximately 70 to 80 % of the water has drained off.

On three occasions a simple material balance was made. The results are shown in Table 5.

TABLE 5. MATERIAL BALANCE - REED BED 1

Date	No. of Days (1)	Amount of Sludge Deliv. mm (2)	Precip-itation mm (3)	Filtrate mm (4)	Diff-erence mm/d (2+3 -4)/1
Feb 22 to March 3	13	103	4	90	1.3
May 21 to 28	8	103	8	70	5.1
June 18 to 26	8	103	36	95	5.5

The difference between the sludge and precipitation entering, and the filtrate leaving the reed bed, was assumed to be lost mainly by evapotranspiration. Other losses due to storage in the plant cells or in the bed drainage media were

considered negligible since most measurements were taken either when the reeds were dormant or had reached their maximum height.

The low losses measured in February and March 1979 indicate that water loss is mainly through evaporation, whereas the higher losses recorded in May and in June identify an additional water loss that we consider to be mainly due to transpiration. The measured evapotranspiration rate lies within the range of 5 to 8 mm/d considered normal for Phragmites communis in this part of Europe (10).

Pedological Analyses of the Sludge Residue

Since the solids residue remains for several years in the rhizosphere of Phragmites communis, it was also necessary to assess its suitability for land application and to determine permissible rates of application. Three sludge residue profiles were made in the first reed bed and bed samples were taken from each profile at three different depths. Afterwards monthly sludge residue samples were taken near these profiles in the upper 2 cm of the solids layer where the mat of rhizomes was the densest. The analyses of these samples are shown in Table 6 and the sludge profiles shown in Figure 2.

The pedological investigations showed that in the dense mat of rhizomes the effective organic substances fluctuate between 40 and 55 % and consist mainly of a stabile humus-like material. At lower depths the organic material content is reduced to 17 and 26 % indicating that the sludge residue is eventually decomposed. The high dehydrogenase activity and the high carbon dioxide efflux are additional indications that the microbial activity in the rhizosphere of Phragmites communis is very active. As expected an increase in this activity between March and June could be measured.

The sludge profiles show that reduction is not taking place. With increasing sludge residue depth the colour of the residue becomes lighter and indicates, as mentioned previously, that with increasing depth the effective organic substances decrease.

Most of the rhizomes and shoots are found in the upper 5 cm of the sludge residue horizon as well as between the interface of the top layer of the bed drainage media and the decomposed sludge. Isolated roots penetrate the gravel layer but this is atypical since Phragmites communis prefers to take root in a finely textured to rolling sandy soil (3,4).

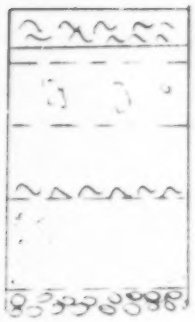
To verify the above findings the filtrate from the first reed bed was analyzed and showed a high soluble salts content which is indicative of the mineralization of organic matter taking place in the rhizosphere. High oxygen levels support the pedological findings that reduction is virtually absent. The high COD : BOD ratio shows that the residues still present in the filtrate are difficult to decompose.

The filtrate analyses taken between 1974 and 1977 as well as those taken in 1979 are shown in Table 7.

TABLE 6. RESULTS OF THE PEDOLOGICAL ANALYSES - REED BED 1

Item	Depth								
	0 - 2 cm			13 - 15 cm			20 - 24 cm		
	min	max	avg	min	max	avg	min	max	avg
pH, ($CaCl_2$)	5.8	6.6	6.1	5.5	6.1	5.8	5.9	5.9	5.9
$CaCO_3$, %	0.4	1.9	1.0	0.0	0.0	0.0	0.0	0.0	0.0
Salt Content, mg/100g	780	1590	1109	600	798	688	630	999	841
Org. Subst. %	38.2	55.5	46.0	39.6	43.8	41.5	17.5	26.0	20.4
C/N, %	8	10	8.5	9	10	9.7	6	12	10
tot. N, %	2.7	3.7	3.2	2.2	2.8	2.5	0.8	1.8	1.3
P_2O_5, mg/100g dry matter	740	880	802	-	-	-	-	-	-
K_2O, "	99	155	1311	77	97	85	45	74	59
MgO, "	93	120	106	110	110	110	68	85	75
Na , "	108	209	152	165	200	183	131	147	138
Fe , "	520	750	636	360	590	507	350	490	407
Dehydrogenase, % Abs.	0.79	2.10	1.55	0.15	0.20	0.17	0.10	0.15	0.12
CO_2, mg/100g dry matter	-	-	500	-	-	-	-	-	-

Profile 1b



0 to 4/5 cm dense mat of roots and rhizomes

4/5 to 6/8 black zone

6/8 to 14/17 lighter zone, single black patches, many earthworms

14/17 to 24/25 dark brown zone, numerous roots and rhizomes

24/25 to 38/40 reddish sand, intermeshed with roots

below 38/40 cm rough gravel, few roots

Profile 1d

0 to 5 cm extensive rhizome mat and roots

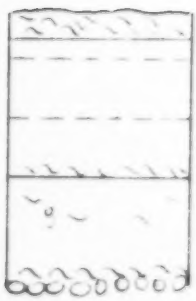
5 to 7 black zone

7 to 17 light brown zone, intermeshed with roots, single black patches, many earthworms

17 to 25/26 dark brown zone, many roots at the bottom

25/26 to 40/42 cm reddish sand, sparsely intermeshed with roots

Profile 1f



0 to 4/5 cm a well developed rhizome and root system

4/5 to 6/7 black zone, well intermeshed with roots

6/7 to 17/18 light brown zone, many earthworms

17/18 to 23/24 dark brown zone, many roots at the bottom

23/24 to 38/40 cm reddish sand, many roots at the sand-gravel interface

FIGURE 2. SLUDGE RESIDUE PROFILES IN REED BED 1

TABLE 7. FILTRATE ANALYSES FROM REED BED 1

Parameters	1974 - 1977 (a)			1977 (b)	1979 (c)
	min	max	avg		
Appearance				yellowish	yellowish
Odour				earthy	musty
Turbidity					35
SS, mg/l					10
Temperature, ^{o}C					8
pH	6.8	8.4	7.6	6.95	7.2
Spec. Cond. mS/cm				1400	1650
COD, mg/l				112	120
BOD_5, mg/l	2	48	11.4	13	6
Free Ammonia, mg/l				4.28	1.83
Nitrites, mg/l				-	0.26
Nitrates, mg/l				102	114
Orthophosphates, mg/l				34.5	180
DO level, mg/l	4.3	13.0	6.9	4.9	11.5

(a) Time Proportional Composite Sample, March 28, 1976, 9-15h
(b) " " " " , Aug 30, 1977, 9-14h
(c) 53 Samples

An analysis of the primary nutrient and heavy metal content of the excess sludge and a fertilizer assessment were carried out by the State Research Institute of Augustenberg. This Institute concluded that the excess sludge generated by the WWT facilities in Ustersbach is not typical. The influence of the brewery waste on the sludge quality could be measured. Compared with other domestic sludges its nutrient and heavy metal content is very low. If used as a fertilizer 94 tons/hectare/annum, i.e. 116 cu.m/hectare/annum of wet sludge could be applied on land. These results are shown in Tables 8 and 9.

Because of the influence of the brewery on the sludge quality parasitological investigations were not carried out.

COMMENTS

Based on what we have observed to date the following comments can be made:

1. Reed beds, i.e. drying beds planted with Phragmites communis, dewater and biologically treat excess stabilized sludge without extraneous energy, chemicals or machinery.

2. Reed beds are able to bring about a higher volume reduction of the sludge than conventional drying beds. Using similar loading rates the sludge level in the conventional bed was approximately 290 % higher than in the reed beds.

3. None of the experiments carried out in 1979 attempted to load the reed beds to failure, i.e. when the beds would have to be emptied every 6 to 12 months. Some of the older records indicate that with the type of sludge available failure will occur at loading rates of approximately 80 to 85 kg/sq.m/a. The most recent experiments sought a loading rate that would ensure a bed turnover of at least 4 to 5 years. A loading rate of 50 kg/sq.m/a should meet this request. It is equivalent to about 0.32 to 0.35 sq.m per capita and lies between the 0.4 sq.m/cap. suggested for conventional beds in Austria and the 0.2 sq.m/cap for German drying beds. The advantage of reed beds actually lies in the low bed turnover rate, e.g. once every 5 years compared with conventional bed turnover rates of once every 1 to 2 months (12,13,14).

4. The reed beds of Ustersbach have been operating on the average 8 to 9 months per year for the last eight years. These beds never experienced a reduction in the dewatering rate or a clogging of the bed drainage media or any other operational difficulties when the above mentioned permissible loading rates were not exceeded.

5. Although a plant assessment was not carried out, harmful

TABLE 8. HEAVY METAL CONTENT IN THE USTERSBACH SLUDGE

Constituents	Amount Measured ppm	Max. Allow. Concentration in Sludge, ppm	Max. Allow. Concentration in Soil, ppm
Pb	30	1200	100
B	4.86	-	-
Cd	1	20	3
Cr	12	1200	100
Cu	60	1200	100
Ni	1	200	50
Hg	-	25	2
Zn	45	3000	300

TABLE 9. FERTILIZER EVALUATION

Item	Amount in one ton of sludge, kg	Fertilizer Type	Equivalent Fertilizer Quantity, kg	Average Fert. Application Rate, kg/100m^2
N	0.74	N-Fert. 25%	2.96	7
P_2O_5	1.02	P-Fert. 16%	6.38	6
K_2O	0.29	K-Fert. 50%	0.58	6
Ca	0.4	Calciumox de	0.56	15
B	0.01	Borax	-	0.2
Cu	0.01	$CuSO_4$	-	0.3

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B	0.01	Borax	-	0.2
Cu	0.01	$CuSO_4$	-	0.3

infections due to parasites or a physical change in the plants due to an overabundance of primary nutrients (11) or heavy metals were never observed. The response of the reeds to the operating conditions was always good. The plant density varied between 136 and 260 reed stems/sq.m and had an average density of 197 stems/sq.m. The maximum height of the reeds varied between 2.5 and 3 m.

6. It has been suggested (14,15) that glass enclosures of the greenhouse variety, or simple roofs to protect the sludge from extraneous moisture could increase the number of dryings per year by 30 to 100 %. Experiments carried out in Karlsruhe (8,9) showed that covered reed beds could achieve similar higher loading rates. By means of a simple roof these reed beds were operating at loadings of 97 kg/sq.m/a without experiencing any difficulties.

7. Experience from the most recent installations, i.e. Neumarkt and Dornstadt, show that the sludge brought to the reed beds does not require any prethickening and/or chemical conditioning. Sludge can be withdrawn either from the return sludge line or directly from the aeration basin(s).

8. After the excess sludge is deposited in the reed beds some of the constituents are removed to produce plant biomass, i.e. roots and rhizomes below the sludge surface, and shoots, stems and leaves above it. Harvesting of the biomass above the sludge surface is an important factor in maintaining a long bed turnover time. The time and extent of the harvesting will depend on the location of the facilities and on the physiology of the reeds.

9. Most of the sludge remains at the site of deposition where it is slowly mineralized and humified. Before this residue is removed tests should be carried out to determine its suitability as a fertilizer supplement. After removal this material should be composted before it is applied on land. At present research is being carried out to assess the windrow method of composting.

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FUNCTIONAL DESIGN OF ACTIVATED SLUDGE PROCESSES

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ABSTRACT

As the amount and type of materials in wastewater increases, there will be a growing need for more finely tuned operational control of wastewater treatment plants. It seems appropriate for designers to consider somewhat more sophisticated modelling practices than those presently employed and to pay more attention to the operational aspects of the plants they design. In this report, a brief overview of a design and operational model for normal wastes and for waste containing toxic organic substrates is given.

The model is used to predict effluent quality and sludge production; these are governed by waste strength, the biokinetic constants and three selectable engineering control parameters. The interrelationships of all of these factors determine the plant efficiency. The engineering model lends itself ideally as a subroutine for economic or costing models because the engineering parameters can be manipulated to optimize design and operation costs for a given degree of treatment. This model is also readily useable by plant personnel to guide operation of the system. A computational example of the use of the model is provided and, in an appendix, a sample computer program in BASIC is provided to facilitate computations on the hand-held minicomputer.

There are several alternative processes for biological treatment of wastewaters, but activated sludge, in one of its several modifications, is the most widely used. Although apparent operational costs may be higher compared to other processes, activated sludge usually compares favorably from the standpoint of initial costs. Higher degrees of purification and flexibility in operation and in design choices are other attributes accorded to these processes. Flexibility is a particularly advantageous attribute because of the general variability of waste flow and composition as well as variability in numerical values for the biokinetic constants used to characterize the biomass.

Today, it is widely recognized that the quality of the effluent with respect to the concentration of soluble organic pollutants, S, remaining after treatment is related to the specific growth rate, μ, of the biomass. The amount of excess biomass, X_v, which is produced is also dependent upon the growth rate. The relationship between μ and S which usually obtains is shown in Figure 1. Equations (1) and (2), which are given in the figure legend, have been found by many investigators to describe the relationship between μ and S for growth of heterogeneous microbial populations on non-inhibitory substrates (Equation 1, Monod relationship) and on inhibitory, or toxic, substrates (Equation 2, the Haldane equation).

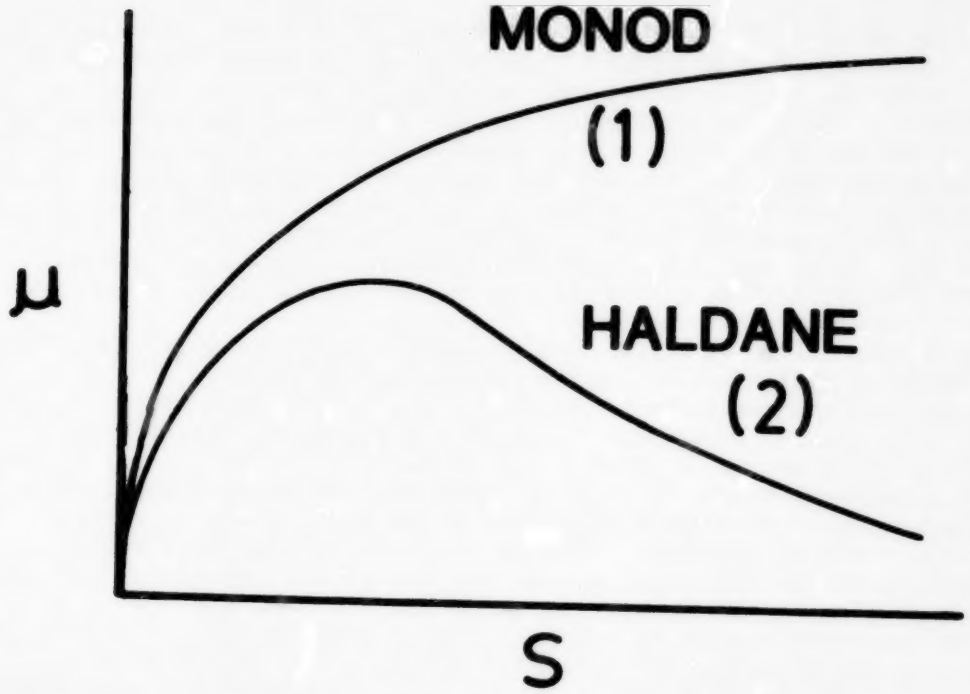

FIGURE 1. EFFECT OF SUBSTRATE CONCENTRATION, S, ON SPECIFIC GROWTH RATE, μ.

$$\mu = \frac{\mu_{max}\,(S)}{K_s + S} \qquad (1)$$

$$\mu = \frac{\mu_{max}\,(S)}{K_s + S - S^2/K_i} \qquad (2)$$

These relationships contain three biokinetic constants which are characteristic of the cell-substrate system: the maximum specific growth rate, μ_{max}, the saturation constant, K_s, and the inhibition constant, K_i. The amount of biomass, X, produced from a given amount of substrate is another biokinetic constant of importance; this constant is termed the "true" cell yield, Y_t. However, it is generally found at slow growth rates, which are maintained in activated sludge systems in order to minimize S in the effluent, that cell decay can occur concurrently with growth; this reduces the cell yield, so that the observed yield, Y_0, is always lower than Y_t. Observed yield, Y_0, is related to Y_t by Equation (3):

$$Y_0 = \frac{Y_t\,\mu_n}{\mu_n + k_d} \qquad (3)$$

In equation (3), a fifth biokinetic constant, k_d, the decay constant, is utilized. The net specific growth rate, μ_n, is computed by subtracting k_d

from the specific growth rate, μ.

$$\mu_n = \mu - k_d \tag{4}$$

The mean cell residence time, θ_c, is defined as the reciprocal of the net specific growth rate

$$\frac{1}{\theta_c} = \mu_n = \mu - k_d \tag{5}$$

It can be seen from equations (1) through (5) that one should seek to control the growth rate in order to control S and the amount of sludge or biomass produced. Consequently, this is the major strategy employed in current design and operational practice. The problem facing designers and operating engineers is how to utilize these concepts optimally in engineering practice. One may ask what engineering controls are available and how can those controls be related to the biokinetic theory to yield predictive model equations for both design and operation. It should be evident that design and operation are inseparable considerations in that the term "flexibility" is meaningless in an engineering sense if they are treated as separate considerations.

The incorporation of engineering parameters and biokinetic constants into conceptual design and operational models can be facilitated with the aid of the flow diagram shown in Figure 2. The waste concentration, S_i, and the flow rate, F, constitute the waste loading to the aerobic reactor, R_1, the operating volume of which is designated as V. This reactor is assumed to be completely mixed and all substrate removal is assumed to take place in R_1. Only solids separation takes place in the clarifier; since it is not 100% efficient, there is an amount of suspended solids, X_e, which escapes in the effluent. All current design and operational models deal only with the soluble substrate, S, in the effluent, i.e., there is currently no way to predict X_e. However, operational guidelines and engineering experience can be applied to keep it as low as possible. Some of the clarifier underflow is wasted; this, along with the cells lost in the effluent, comprises the mass of excess sludge wasted daily, X_w. The remaining portion of underflow cells is used for recycle and can be adjusted to a certain return sludge concentration, X_R. These cells are pumped to R_1 at a flow rate αF; α is the recycle flow ratio, F_R/F. X_R and α are two important engineering variables which can be used to control the growth rate, μ_n, or its reciprocal, θ_c, and thereby S and X_w as shall be shown. The return sludge dosing tank, R_2, shown in dashed lines represents an option which allows the operator to control X_R rather than recycling at the variable concentration delivered by the clarifier. The symbol S_R represents the concentration of substrate in the recycle flow. In design calculations S_R can be assumed to have a specific concentration or it can be set equal to S in the reactor effluent or it may be assumed to be negligible, i.e., $S_R = 0$.

Another important engineering design factor is the mean hydraulic detention time in the reactor, $\bar{t}$ and its reciprocal, D, the dilution rate.

$$\bar{t} = V/F = 1/D \tag{6}$$

D, or $\bar{t}$, α and X_R are the essential control parameters to be considered in design and operation.

DEVELOPMENT OF MODEL EQUATIONS

For purposes of illustration, let us assume that S_i is non-inhibitory so that equation (1) applies. We are interested in describing the

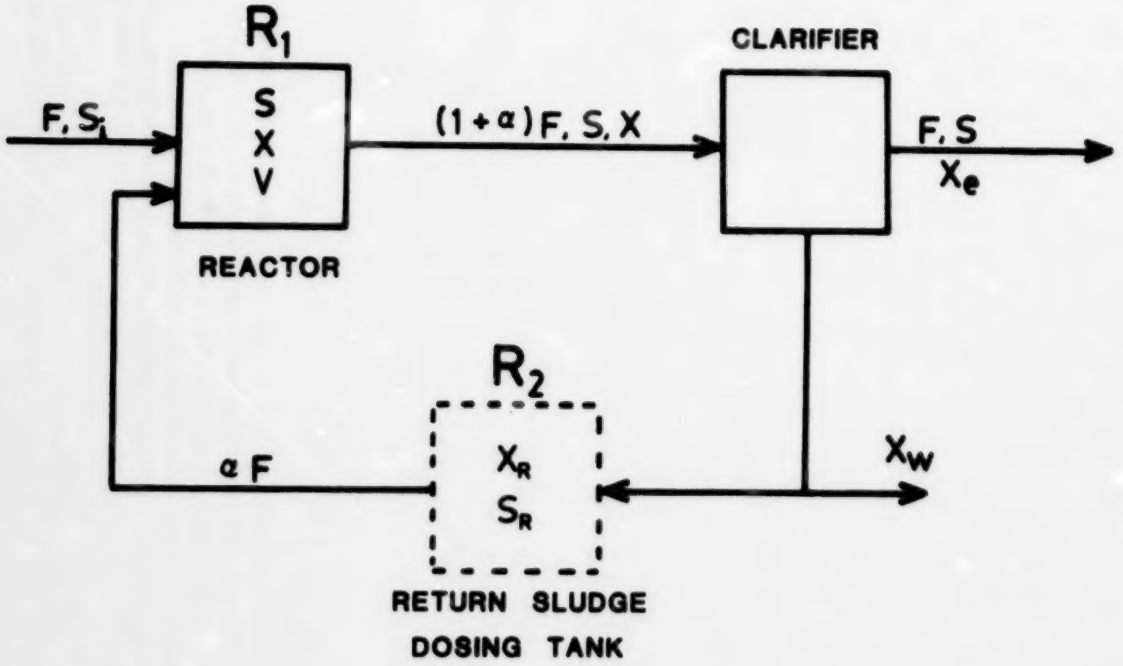

FIGURE 2. FLOW DIAGRAM OF AN ACTIVATED SLUDGE PROCESS WITH CONSTANT CONCENTRATION OF RECYCLE SLUDGE, X_R

biochemical purification process which occurs in R_1. The changes in S and X taking place in the reactor can be depicted by writing a materials balance for R_1 as follows:

	INPUT	REACTION	OUTPUT	
$V(dX/dt) =$	αFX_R	$+\mu XV - k_d XV$	$-(1+\alpha)FX$	(7)
$V(dS/dt) =$	$FS_i + \alpha FS_R$	$-XV\mu/Y_t$	$-(1+\alpha)FS$	(8)

The assumption that a pseudo-steady state will occur in heterogeneous biomass systems has been shown to be correct provided R_1 is completely mixed with respect to S and X, and provided the inputs S_i, F, α and X_R are held relatively steady.(1)(2)(3)(4)(5) These conditions are subject to engineering control, and thus one can set the rate of change in X and S to 0. By doing this, letting F/V equal the dilution rate, D, and substituting equation (1) for μ, predictive equations for S and X can be obtained:

$$X = \frac{Y_t[S_i + \alpha S_R - (1+\alpha)S] + \alpha X_R}{1 + \alpha + k_d/D} \tag{9}$$

$$S = \frac{-b \pm (b^2 - 4ac)^{1/2}}{2a} \tag{10}$$

$$a = \left[[(1+\alpha)D + k_d] - \mu_{max}\right](1+\alpha)$$

$$b = \mu_{max}\left[S_i + S_R + \frac{\alpha X_R}{Y_t}\right] + [(1+\alpha)D + k_d][(1+\alpha)K_s - (S_i + \alpha S_R)]$$

$$c = -(S_i + \alpha S_R)[(1+\alpha)D + k_d]K_s$$

In the above solution, S_R was assumed to have a specific value. (See Appendix 1 for equations in which it is assumed to be equal to S. Appendix 1 also gives equations for inhibitory substrates.) A materials balance may be written around the clarifier, yielding the following equation for X_w:

$$X_w = D\left[1+\alpha - \frac{\alpha X_R}{X}\right]XV = \mu_n XV = \left[1+\alpha-\alpha\frac{X_R}{X}\right]XF \tag{11}$$

Equations (9), (10) and (11) provide the designer with the relationships needed to make design decisions and they also provide the operating engineer with analytical tools to assess the effects of changes in loading; i.e., S_i and/or F, or in biokinetic constants or in the engineering parameters D (or t), X_R and α. The latter three are the only engineering control measures available unless S_i and F are "equalized" at the plant site or unless certain values of S_i and F are proscribed in an overall wastewater systems management program.

USE OF THE DESIGN MODEL

Equations (9), (10) and (11) are very useful in design and analysis and we will illustrate this with a sample design problem. However, there are two aspects of the model which deserve some comment before proceeding with the example problem.

First, it is apparent that solution of the equations for X and S for various values of the constants and control parameters could be laborious if done by hand or by calculator, and it is therefore ideal to employ one of the programmable hand-held minicomputers which have recently begun to replace calculators. A program in BASIC is given in Appendix 2. This program calculates S, X, X_w and μ_n (or θ_c); it puts tremendous computational power at the fingertips of the designer or analyst and permits one to make a variety of computations which might be neglected in the functional design stage if this computational power were not so readily available.

The other necessary consideration is the determination of representative values for the biokinetic constants. These are usually determined during treatability studies, and methods for their determination are described elsewhere.(2)(3)(6)(7)(8) Average values gleaned from the literature (9)(10) (see Appendix 3) may be used, but it is recommended that the values be obtained by experiment on the waste in question.

It is important to emphasize that design models cannot predict adequately unless the wastewater and the biomass kinetic constants are adequately characterized. This should be part of the treatability and functional design studies. For example, it is essential that batch growth rate studies be made to determine the shape of the curve relating μ and S (see Figure 1). This permits one to select the applicable model equations given in Appendix 1.

To illustrate the use of the model in design and/or operation, let us consider a non-inhibitory waste with the design inputs given in Table 1.

TABLE 1
CHARACTERISTICS OF WASTEWATER AND BIOMASS

F	20,000 m^3/day (5.3 MGD)
Si*	500 mg/l, range 700 - 300 mg/l
μ_{max}	0.38 hr^{-1}
K_s	100 mg/l
Y_t	0.6 mg/mg
k_d	0.0066 hr^{-1}

*S_i may be given in terms of soluble BOD, ΔCOD or ΔTOC

One way to examine the many options available to an engineer is to construct graphs showing the effect of hydraulic detention time ($\bar{t} = V/F$) or dilution rate ($D = F/V$) on effluent quality, S, and reactor biomass, X, for the range of waste strengths, S_i, expected. As a beginning one might employ values of α and X_R commonly used or anticipated; e.g., $\alpha = 0.25$, $X_R = 10{,}000$ mg/l.

Design for Effluent Quality

The program given in Appendix 2 can be entered into a pocket computer; it is executed by entering the characteristics given in Table 1, and the selected values of α and X_R for various values of D; i.e., $1/\bar{t}$. A plot of the calculated values of reactor biomass concentration, X, and soluble substrate concentration, S, is shown in Figure 3.

These graphs show the effect of the dilution rate, D, or hydraulic retention time, $\bar{t}$, on S and X over the range of S_i values expected. The biomass, or mixed liquor suspended solids concentration, X, will not vary much because it is governed primarily by the recycle sludge concentration,

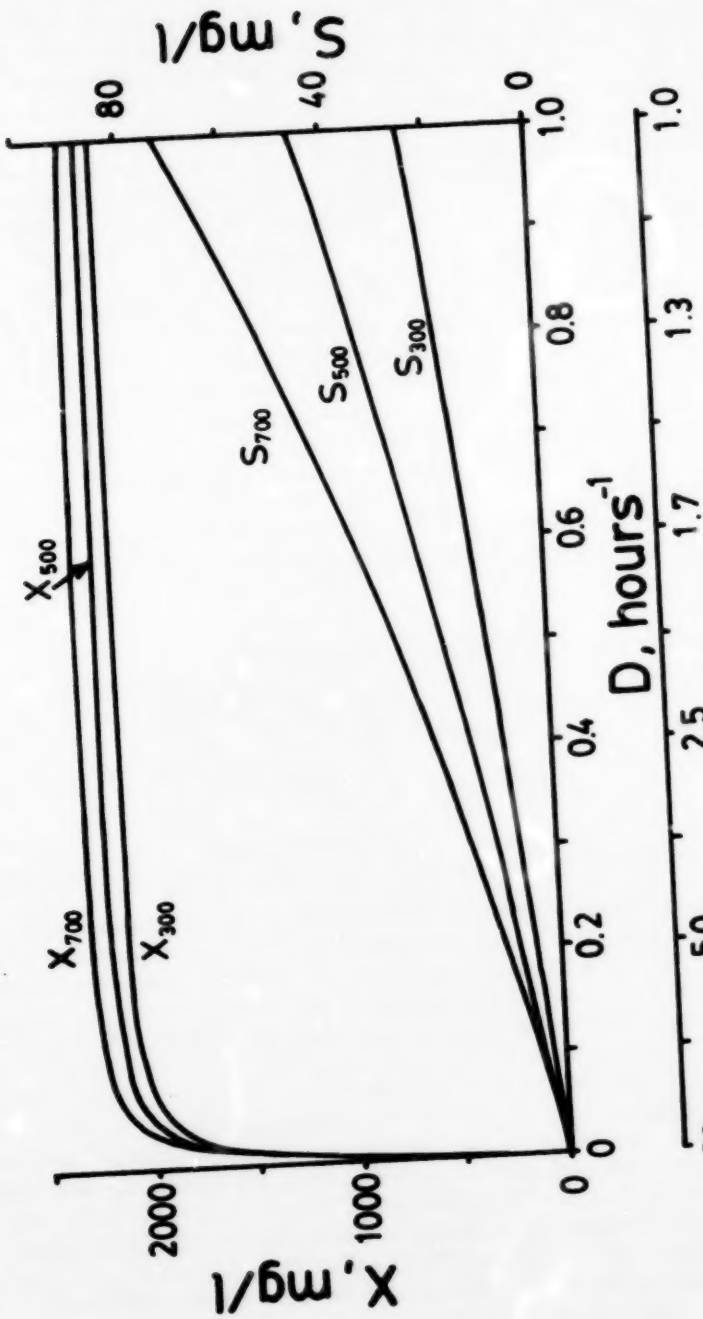

FIGURE 3. EFFECT OF MEAN HYDRAULIC RETENTION TIME, $\bar{t}$, ON X AND S AT FEED CONCENTRATIONS OF 300, 500 AND 700 mg/l
$\alpha = 0.25$; $X_R = 10,000$ mg/l

X_R, of 10,000 mg/l, rather than by growth at these levels of S_i. It is seen that as the dilution rate increases, or mean hydraulic detention time decreases, greater amounts of soluble substrate exit the reactor. It is also evident that soluble substrate leaving the reactor increases with increasing S_i.

Assume that effluent requirements are such that S cannot be greater than 10 mg/l. This limitation can be met when S_i = 700 mg/l if $\bar{t}$ = 5 hours. Thus, the volume of reactor can be determined. If it were known that S_i would run at 700 mg/l for a relatively small portion of the time, one could design for the average S_i of 500 mg/l with $\bar{t}$ = 4 hours, thus reducing the volume of reactor R_1 at the risk of permitting S to be higher than 10 mg/l a portion of the time. A designer may wish to employ a 4-hour detention time and hold S to 10 mg/l during the expected period of high S_i, by providing for increased α and/or X_R. In this case the costs of pumping and controlling X_R are balanced against the costs of construction for R_1. The effect of varying X_R at the average S_i of 500 mg/l is shown in Figure 4. It is seen that at $\bar{t}$ = 4 hours, an increase in X_R to 15,000 mg/l provides an S well below 10 mg/l at S_i = 500 mg/l, whereas at S_i = 700 mg/l, S would be 9 mg/l.

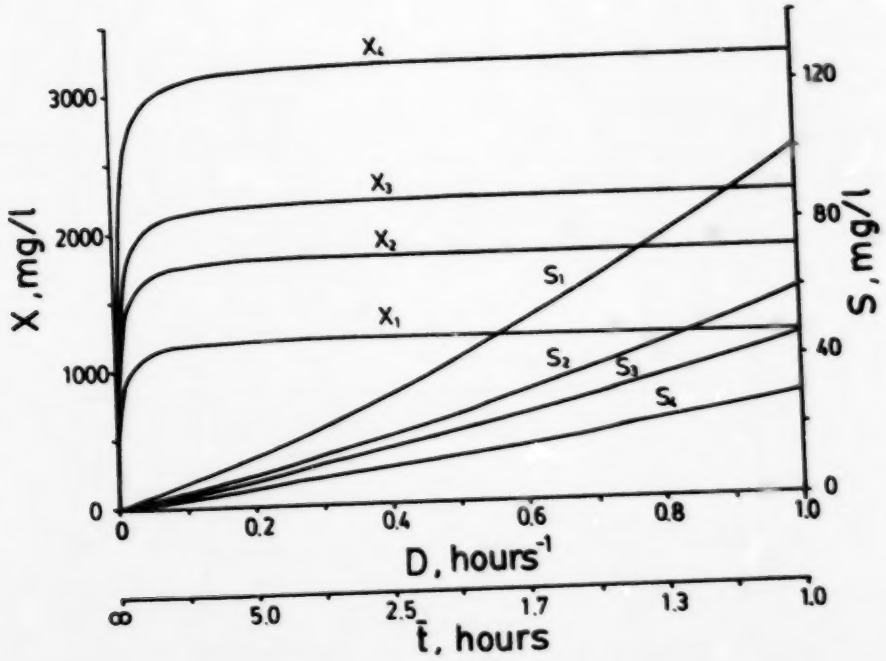

FIGURE 4. EFFECT OF MEAN HYDRAULIC RETENTION TIME, $\bar{t}$, ON X AND S AT VARIOUS RECYCLE SLUDGE CONCENTRATIONS.
X_{R1} = 5,000 mg/l; X_{R2} = 8,000 mg/l; X_{R3} = 10,000 mg/l;
X_{R4} = 15,000 mg/l; α = 0.25; S_i = 500 mg/l

The lower construction cost for a reactor of smaller volume is exchanged for the expense of providing a higher X_R. Whether 15,000 mg/l return sludge concentration could be provided without some sludge thickening would depend upon the settling and compacting characteristics of the biomass as well as the clarifier design. Often, thickening is employed prior to treatment and disposal of the sludge which could give one the option of using a portion of the thickened sludge for recycle to R_1 in order to increase the concentration of X_R. Although increased X_R values result in lower aeration tank volumes, they also result in higher values for mixed liquor suspended solids, X, which necessitates increased costs for aeration and mixing. It is apparent that the model can be utilized as part of a cost-effective analysis to determine an optimal economic balance between aeration basin volume and mixing and aeration costs.

Figure 5 illustrates the effect of varying α on effluent substrate concentration. X_R is held constant at 10,000 mg/l. As was the case with increases in X_R (see Figure 4), an increase in α results in better reactor performance with respect to biochemical purification. Consequently, provided that the required X_R levels are attainable, increased α values result in lower aeration tank volumes. One can therefore perform an analysis that evaluates the economics of varying α by utilizing the design

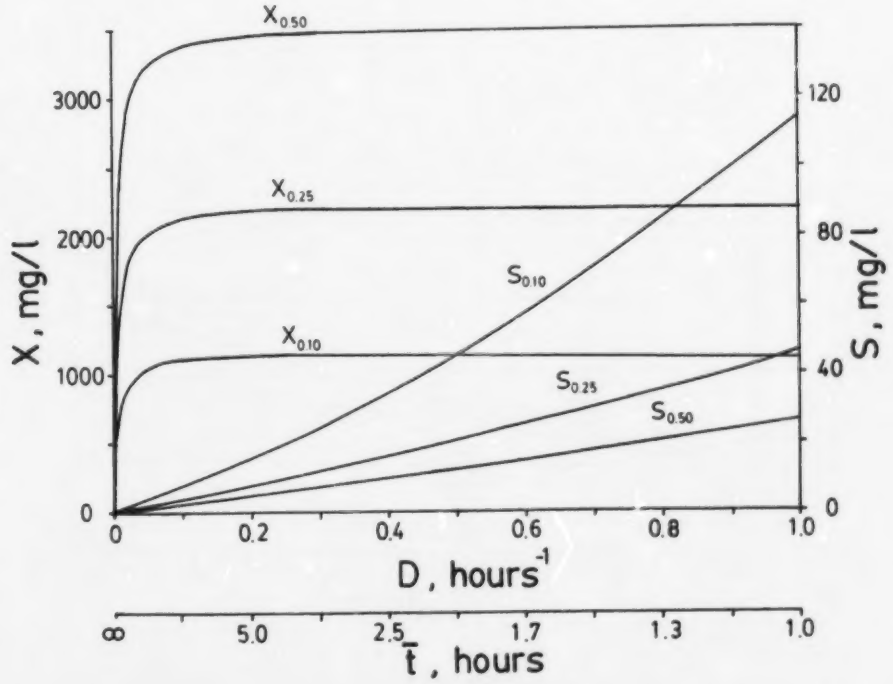

FIGURE 5. EFFECT OF MEAN HYDRAULIC RETENTION TIME, $\bar{t}$, ON X AND S AT RECYCLE FLOW RATIOS OF 0.10, 0.25 and 0.50.
X_R = 10,000 mg/l; S_i = 500 mg/l

model in a cost-effective algorithm. In the case of increased α, the savings in construction costs for decreased reactor volume may be offset by increased pumping costs in addition to increased requirements for aeration and mixing. Settling tank costs will also increase because of the greater solids loading exerted on the secondary clarifier due to increases in α.

Design for Minimal Sludge Production

There is yet another aspect involving engineering and economic trade-offs which affects the design and operation of the plant. The by-product of treatment is sludge in excess of that needed for recycle. Sludge disposal is a costly matter and a decision must be made as to whether it is better to produce, treat and dispose of whatever sludge is developed at the μ_n (or θ_c) selected to meet effluent standards or to control growth rate in order to minimize sludge production.

Previously, we have seen that the required S of 10 mg/l could be obtained with $\bar{t}$ = 5 hours, α = 0.25 and X_R = 10,000 mg/l. These are reasonable values and they lead to a net specific growth rate of 0.50 $days^{-1}$; i.e., θ_c of 2.0 days. In accordance with equations (1) and (4), the growth rate cannot be greater than this value or S will be above that required by the design. However, an even lower growth rate would lower S even further and, in accord with equation (11), would also reduce the amount of excess sludge. For the design flow and biokinetic constants given, the growth rate can be reduced by increasing α (a greater pumping rate) or X_R (thickening of the sludge beyond 10,000 mg/l)or $\bar{t}$ (larger volume of reactor). For example, consider Figure 6 which shows, for α = 0.25, the relationship between X_w and D (or $\bar{t}$) at several values for X_R. The curve labelled X_R = 10,000 mg/l shows the reduction of X_w with increased $\bar{t}$. The plotted triangle highlights the value of X_w at $\bar{t}$ = 5 hours, the value previously selected to meet the effluent requirement of 10 mg/l. At a $\bar{t}$ of 23 hours (D = 0.044 $hours^{-1}$), the curve crosses the x-axis, so that 23 hours is the calculated $\bar{t}$ for "total oxidation", an extended aeration process design. It should be emphasized that, as the curve falls away to the abscissa, both X_w and S decrease because μ_n decreases. The excess sludge, X_w, can be reduced with less increase in reactor tank volume if a thicker return sludge is provided. For example, at X_R = 15,000 mg/l, $\bar{t}$ for X_w = 0 is just over 15 hours, whereas at X_R = 20,000 mg/l, $\bar{t}$ for total oxidation is approximately 11 hours.

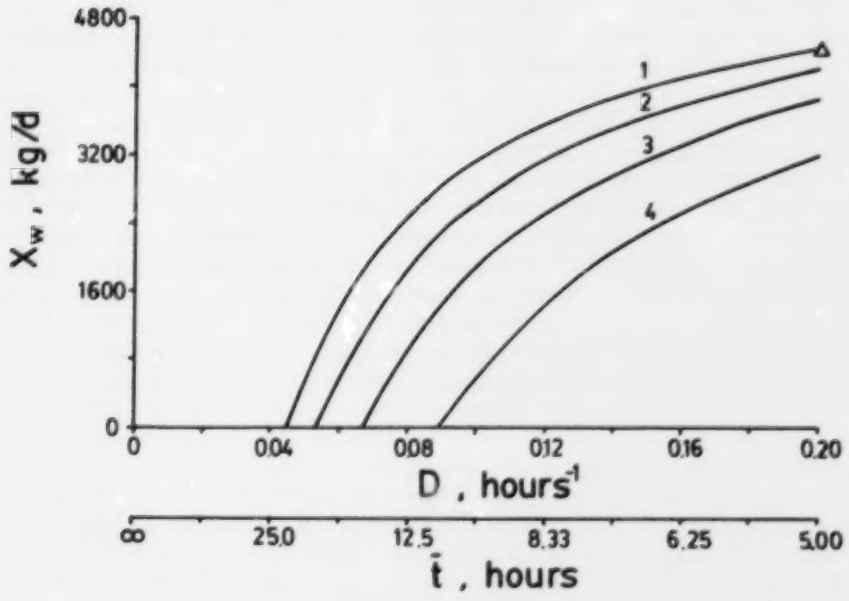

FIGURE 6. EFFECT OF MEAN HYDRAULIC RETENTION TIME, $\bar{t}$, AND RECYCLE SLUDGE CONCENTRATION, X_R, ON PRODUCTION OF EXCESS SLUDGE, X_w. $\alpha = 0.25$; $S_i = 500$ mg/1; curve 1, $X_R = 10{,}000$ mg/1; curve 2, $X_R = 12{,}000$ mg/1; curve 3, $X_R = 15{,}000$ mg/1; curve 4, $X_R = 20{,}000$ mg/1. All curves provide lower S values than the required effluent concentration of 10 mg/1.

Excess sludge production can also be reduced by increasing α as seen in Figure 7, in which X_R is held at 10,000 mg/1. The curve labelled $\alpha = 0.25$ is the same as the curve for $X_R = 10{,}000$ mg/1 in Figure 6 and the plotted triangle has the same significance as before. At an α of 0.25, the $\bar{t}$ for zero sludge production is just over 13.5 hours, whereas it is reduced to 9 hours at $\alpha = 1.0$. However, it should be mentioned that high recycle ratios tend to reduce the concentration of sludge in the clarifier underflow so that it may not be possible to reliably attain an X_R of 10,000 mg/1. Some provision for thickening sludge may be needed. In this case, if one chose to control X_R using a sludge dosing tank, as suggested in Figure 2, the thickener could serve as a source of sludge with which to load the dosing reactor.

SUMMARY AND CONCLUSIONS

The mathematical model herein presented for functional design and analysis of activated sludge processes is founded on principles first delineated by microbial kineticists to describe growth and substrate removal in rather defined and highly controlled systems, i.e., pure cultures with defined carbon sources. Many years of research effort have gone into

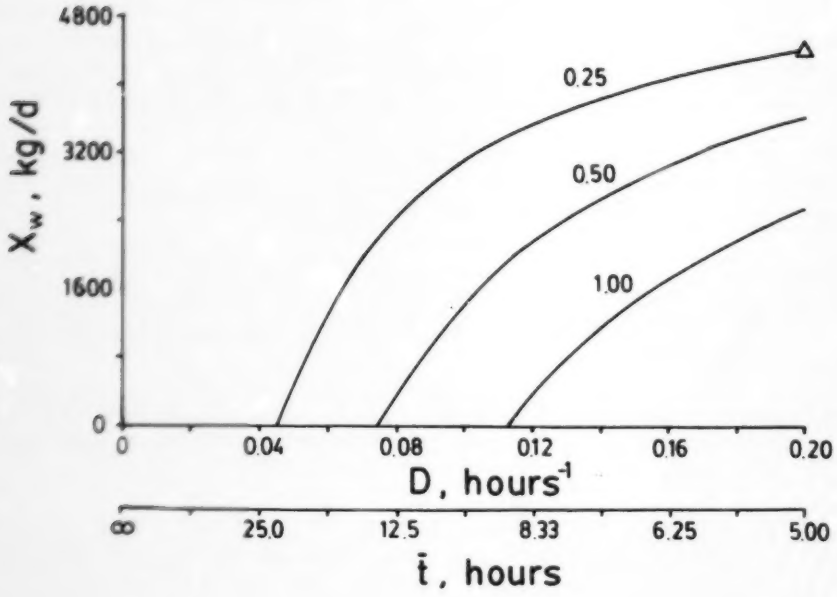

FIGURE 7. EFFECT OF MEAN HYDRAULIC RETENTION TIME, $\bar{t}$, AND RECYCLE FLOW RATIO, α, ON PRODUCTION OF EXCESS SLUDGE, X_w. X_R = 10,000 mg/l; S_i = 500 mg/l; recycle flow ratios (indicated on curves) are 0.25, 0.50 and 1.0. All curves provide lower S values than the required effluent concentration of 10 mg/l.

developing and modifying the model for the less well-defined systems encountered in the water pollution control field. These principles are now well recognized and it should be pointed out that there is need for continuing work in order to provide fine-tuning of performance and broadening of the scope of application for this process. The major engineering controls now available to the designer are mean hydraulic retention time, $\bar{t}$, recycle flow ratio, α, and recycle cell concentration, X_R. These engineering controls, along with inflowing substrate concentration, S_i, determine the specific growth rate which, in turn, largely defines the concentration of soluble substrate in the reactor effluent and the amount of excess sludge which will be produced during treatment. The model provides the designer and systems analyst with a means to assess the effect of physical changes he can impose on the important process outputs; thus, the model can be a very useful tool in making decisions. It is important to note, however, that the model contains, in addition to the variables which are subject to engineering control, five biokinetic constants which are a function largely of the biomass and the nature of the substrate. The model is of little use unless this interactive character has been examined. Characterization of wastewaters and the biomass must precede optimal use of the model. This is the most costly and important part of functional design which, if neglected, reduces the usefulness of this or any other model.

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APPENDIX I

TYPICAL RANGES OF BIOKINETIC CONSTANTS

Constant	Range	Units
Y_t	0.35 - 0.65	mg biomass/mg COD
k_d	0.003 - 0.010	hr^{-1}
μ_{max}	0.20 - 0.60	hr^{-1}
K_s	40 - 200	mg/ℓ (COD)
K_i	60 - 500	mg/ℓ (COD)

APPENDIX II
STEADY STATE PREDICTIVE EQUATIONS FOR ACTIVATED SLUDGE WITH CONSTANT X_R

NONINHIBITORY SUBSTRATE (Monod)

CASE A: $S_R = S_R$

$$S = \frac{-b \pm (b^2 - 4ac)^{1/2}}{2a}$$

$$a = \left[\left[(1 + \alpha)D + k_d\right] - \mu_{max}\right](1 + \alpha)$$

$$b = \mu_{max}\left[S_i + \alpha S_R + \frac{\alpha X_R}{Y_t}\right] + \left[(1 + \alpha)D + k_d\right]\left[(1 + \alpha)K_s - (S_i + \alpha S_R)\right]$$

$$c = -(S_i + \alpha S_R)\left[(1 + \alpha)D + k_d\right]K_s$$

$$X = \frac{Y_t\left[S_i + \alpha S_R - (1 + \alpha)S\right]}{1 + \alpha + k_d/D} + \alpha X_R$$

$$X_w = VX\mu_n$$

CASE B: $S_R = S$

$$S = \frac{-b \pm (b^2 - 4ac)^{1/2}}{2a}$$

$$a = \left[(1 + \alpha)D + k_d\right] - \mu_{max}$$

$$b = \mu_{max}\left[S_i + \frac{\alpha X_R}{Y_t}\right] + \left[(1 + \alpha)D + k_d\right](K_s - S_i)$$

$$c = -S_i\left[(1 + \alpha)D + k_d\right]K_s$$

$$X = \frac{Y_t(S_i - S) + \alpha X_R}{1 + \alpha + k_d/D}$$

$$X_w = VX\mu_n$$

APPENDIX II (continued)

INHIBITORY SUBSTRATE (Haldane)

CASE A: $S_R = S_R$

$$S: \quad S^3 + bS^2 + cS + d = 0$$

$$a = \frac{(1 + \alpha)\left[(1 + \alpha)D + k_d\right]}{K_i}$$

$$b = \left[(1 + \alpha)D + k_d\right]\left[1 + \alpha - \frac{(S_i + \alpha S_R)}{K_i}\right] - (1 + \alpha)\mu_{max}$$

$$c = \mu_{max}\left[(S_i + \alpha S_R) + \frac{\alpha X_R}{Y_t}\right] + \left[(1 + \alpha)D + k_d\right]\left[(1 + \alpha)K_s - (S_i + \alpha S_R)\right]$$

$$d = -(S_i + \alpha S_R)\left[(1 + \alpha)D + k_d\right]K_s$$

$$X = \frac{Y_t\left[(S_i + \alpha S_R) - (1 + \alpha)S\right] + \alpha X_R}{1 + \alpha + k_d/D}$$

$$X_w = VX\mu_n$$

CASE B: $S_R = S$

$$S: \quad aS^3 + bS^2 + cS + d = 0$$

$$a = \frac{(1 + \alpha)D + k_d}{K_i}$$

$$b = \left[(1 + \alpha)D + k_d\right](1 - S_i/K_i) - \mu_{max}$$

$$c = \mu_{max}\left[S_i + \frac{\alpha X_R}{Y_t}\right] + {}_t(1 + \alpha)D + k_d\,(K_s - S_i)$$

$$d = -S_i\left[(1 + \alpha)D + k_d\right]K_s$$

$$X = \frac{Y_t(S_i - S) + \alpha X_R}{1 + \alpha + k_d/D}$$

$$X_w = VX\mu_n$$

Appendix III

BASIC Computer Program For Constant X_R Activated Sludge Model for Noninhibitory Substrate

```
 10:  "XR MONOD"
 30:  REM  MONOD/NONINHIB. SUB
 40:  REM  ENTER BIOKINETIC CONSTANTS
 50:  INPUT  "ENTER UMAX", U1
 60:  INPUT  "ENTER Ks", K2
 70:  INPUT  "ENTER YIELD, Yt", Y1
 80:  INPUT  "ENTER kd(hr-1)", K1
 90:  REM  ENTER ENGINEERING CONSTANTS
100:  INPUT  "ENTER DILUTION RATE (HR-1)", D
110:  INPUT  "ENTER REC. SLUDGE CONC., XR", X
120:  INPUT  "ENTER WASTE CONC., Si", S2
130:  INPUT  "ENTER REC. RATIO, ALPHA", A1
140:  REM  REACTOR VOL. FOR Xw PREDICTION
145:  REM  VOLUME IN M^3
150:  INPUT  "ENTER REACT. VOLUME", V1
160:  P = S2+((A1*X)/Y1)
170:  Z = ((1+A1)*D)+K1
180:  REM  SR = SR OR S ?
190:  INPUT  "ENTER 1 IF SR=SR; 2 IF SR=S", S5
200:  ON S5GOTO 210, 280
210:  REM  NEED VALUE FOR SR
220:  INPUT  "ENTER SR", S1
230:  Q = S2+A1*S1
240:  A = (Z-U1)*(1+A1)
250:  B = U1*(P+S1)+Z*((1+A1)*K2-Q)
260:  C =-Q*Z*K2
270:  GOTO 320
280:  REM  SR = S
290:  A = Z - U1
300:  B = U1*P+(Z*(K2-S2))
310:  C =-S2*Z*K2
320:  R = (B^2-4*A*C)^0.5
330:  S3 = (-B+R)/(2*A)
335:  IF S3>=0.0 THEN 350
340:  S3 = (-B-R)/(2.*A)
350:  ON S5GOTO 370, 420
360:  REM  POS. ROOT ONLY
370:  X1 = Y1*(Q-((1+A1)*S3))+A1*X
380:  X1 = X1/(1+A1+K1/D)
410:  GOTO 460
420:  X1 = (Y1*(S2-S3)+A1*X)
```

```
430:  X1 = X1/(1+A1+K1/D)
460:  REM  COMPUTE NET GROWTH RATE, Un
470:  REM  Un IN DAYS-1
480:  U2 = D*(1+A1-(A1*X/X1))*24
500:  REM  COMPUTE WASTE SLUDGE, Xw
510:  REM  Xw in KG/DAY
520:  X3 = V1*X1*U2*0.001
540:  REM  COMPUTE MCRT
550:  M1 = 1/U2
570:  REM  COMPUTE SUB. UTIL. RATE, U (DAYS-1)
580:  U5 = D*(S2-S3)/X1*24
600:  REM  COMPUTE F:M (DAYS-1)
610:  F1 = D*S2/X1*24
630:  REM  ***PRINT OUT VALUES***
650:  LPRINT  " X = "
660:  LPRINT USING "+######.#"; X1
670:  LPRINT  " S = "
680:  LPRINT USING "+####.##"; S3
690:  LPRINT  " U-NET = "
700:  LPRINT USING "+####.##"; U2
710:  LPRINT  " MCRT = "
720:  LPRINT USING "+####.#"; M1
730:  LPRINT  " Xw = "
740:  LPRINT USING "+#####.#"; X3
750:  LPRINT  " U = "
760:  LPRINT USING "+##.##"; U5
770:  LPRINT " F:M "
780:  LPRINT USING "##.##"; F1
790:  END
```

POLLUTION CONTROL
THE HIDDEN PROFIT CENTRE

J.W. Gerald Rupke
Rupke & Associates Ltd.

INTRODUCTION

The enforcement of environmental legislation has frequently been viewed by industry as a costly and unproductive activity only reluctantly persued. In many cases industry has responded to environmental controls only as a direct result of enforcement of regulations by environmental agencies. Unfortunately this view of environmental control is not totally unwarrented. In many cases control systems were designed as an "end of pipe solution" to a problem that originated in the production facility. It is our experience that in most cases, significant savings can be obtained by a rigorous review of in-plant pollution sources and load reduction efforts at the source. These savings come about as a result of decreased capital and operating costs when compound to an end of pipe system.

The subject of this paper is however, not cost reduction but turning pollution control into a money making effort. It is not always possible to show a profit from pollution control, but in many cases the close scrutiny of pollution sources can identify raw and final product losses of such a magnitude that the savings obtained in the production facility more than offset the costs of pollution control.

The potential for savings of this type are not limited to a few isolated industrial sectors. Table I is a partial listing of industris that have high potential for increasing profits through pollution control efforts.

INDUSTRIES WITH HIGH POTENTIAL SAVINGS

TABLE I

INDUSTRIAL SECTOR	POTENTIAL SAVINGS
Food Processing	product loss; saleable wastes; raw material loss
Pulp & Paper	water recycle; pulp recovery; energy savings
Metal Finishing	metal recovery; water recycle; chemical loss
Slaughterhouse	blood and offal recovery; energy savings
Tannery	chromium recovery; chemical loss
Petrochemical	energy savings; product loss
Cheese Production	whey recovery; energy savings; saleable wastes
Brewery	energy savings; saleable wastes

This paper deals specifically with a case history of a food processor operating a 200m^3/day activated sludge system followed by an aerated lagoon and (2)

two stabilization ponds. The environmental regulatory agency required a high quality effluent of 15 mg/L BOD and 5 mg/L Total Phosphorus.

Sporatic poor operation, odour problems from the lagoons and potential litigation led the owners to retain Rupke & Associates Ltd. to deal with the environmental problems.

CASE HISTORY

Rupke & Associates Ltd. were retained in Febraury, 1980 to assist the company to meet it's environmental objectives. When Rupke & Associates Ltd. staff arrived on site immediate action was implemented in order to rectify the high BOD levels (and anerobic conditions) in the stabilization ponds. These ponds were ready to discharge by May, 1980. Starting in September, 1981 to the present Rupke & Associates Ltd. staff have been operating the waste treatment plant on behalf of the company. This is done under a fixed monthly fee operating contract. The following three (3) areas of savings and profit have resulted from this arrangement.

Maximizing Use of Existing Equipment

The activated sludge system installed at this site is a conventional system using one (1) primary clarifier, two (2) mechanically aerated aeration tanks and a single large final clarifier. The aerated lagoon was being used for stabilization and storage of the waste activated sludge, with effleunt polishing being provided by two (2) final stabilization ponds. The primary sludge was being thickened and disposed of at no cost along with other dry food wastes as cattle feed. It was felt by the company that the treatment plant and specifically the aeration system and sludge thickener were being pushed to their maximum capacity. The regulatory agency was concerned about potential odours from the aerated lagoon being used for waste activated sludge stabilization and storage. The professional staff supplied by Rupke & Associates Ltd. during the first year of operation were able through minor piping and operating changes to achieve significantly higher sludge solids from the thickener resulting in a more acceptable cattle feed product as well as increased capacity in the sludge thickener.

The aeration system was evaluated and found to be of sufficient capacity that during periods of lower production a single aeration tank could be used while maintaining effleunt quality. Alteration of operating practises have resulted in a consistently high quality effluent being produced while demonstrating a significant reserve capacity in the system. The company has recently offered to treat a concentrated waste stream as a service to an adjacent company. This would be done at a profit of course.

Thus through the use of professional operators the situation has changed from an overloaded situation to one of selling reserve capacity to a neighbour without making any capital expenditure.

There are many other areas where the use of unskilled or inexperienced operators may result in misleading information being supplied to management. As a result expensive and inappropriate decisions are being made.

The most common errors are made in the areas of flow monitoring, organic load evaluation, sludge disposal needs and chemical consumption requirements. In each instance the use of professional trained staff, although more expensive for manpower, will frequently result in a lower total cost of providing pollution control. As the system size increases the significance of these potential savings increase dramatically.

Raw Product Savings

One of the major concerns of the regulatory agency was the discharge of 15 - 20 mg/L phosphorus in the effluent stream. The conventional approach to phosphorus removal using chemical precipitation was investigated and found to require a capital expenditure of $20,000 and an annual operating cost of $5,000.

Prior to implementing a phosphorus removal program at the pollution control plant a in-plant survey in the production facility was undertaken. It was found that approximately 75% of Total Phosphorus load originated at a single site where the product was being sprayed with a concentrated phosphate solution. Through a program of spray reduction and capture of the overspray it was possible to reduce the influent phosphorus levles to the point that the activated sludge system was capable of incorporating virtually all the phosphorus into the growth of activated sludge.

Once the company recognized the magnitude of savings available in this area tight controls on chemical loss were instituted. The first year savings were in the area of $15,000 to $20,000 on chemical usage. In fact the production facilities control on phosphorus loss became so tight that the wastewater became nutrient deficient in phosphorus. Rather then relax its in-plant control of food grade phosphorus the company chose to purchase less expensive fertilizer grade phosphorus to supplement the nutrient requirements of the pollution control plant.

Waste Product Utilization

The stabilization and storage of waste activated sludge in the off-stream aerated lagoon was frowned on by the environmental regulatroy agency because of the high potential for odour generation, especially during the period of spring turnover. The system had worked reasonably well for in excess of five (5) years, however, springtime loads had been gradually increasing due to the accumulation of solids on the lagoon bottom. It was anticipated that a major odour problem would develop in the near future.

When evaluating the alternatives for handling the waste activated sludge it became evident that significant capital monies would be required. When handled as a waste to be disposed of the minimum capital required was $340,000 with an anticipated annual operating cost of $120,000/a.

Since the pollution control plant did not treat any domestic sewage it was decided to investigate the option of dewatering the waste activated sludge along with the primary sludge and incorporating the resultant sludge cake in the cattle feed. Since the value of the solids as a cattle feed were dependent on the water content (the lower the water content the greater the monitary value) it was imperative to provide some form of dewatering device.

Extensive pilot and full scale testing work was done using both centrifuges and plate and frame devices to evaluate the combine sludges dewatering characteristics. Capital costs varied from a high of $555,000 for a plate and frame system to a low of $225,000 for a used scroll centrifuge installation. The minimum operating cost was $50,000/a for the scroll type centrifuge. The resultant cattle feed will now be sold for approximately $25,000/a.

The centrifuge installation was put into service early October, 1983. Initial operating results show that with careful control the anticipated solids concentration of 25% can be achieved.

What was once a costly disposal problem is now a saleable product.

SUMMARY AND CONCLUSIONS

1. The use of professional operators can maximixe the use of existing equipment and delay or remove the need for capital expenditures.
2. The control of pollutants at source can result in significant raw product cost savings as well as improved effleunt quality.
3. An "end of the pipe solution" is usually not the minimum cost solution.
4. The concept of viewing a waste stream as a resource rather then a problem is practical, resulting in lower capital and operating costs and at times can be turned into a saleable product.
5. This concept is applicable to a wide range of industries.
6. The net effect of pollution control can be a positive one on a company's balance sheet.

SLIME GROWTH ON FINE BUBBLE DIFFUSERS

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INTRODUCTION

In 1980 the Ontario Ministry of the Environment commissioned Gore & Storrie Ltd. to design a 56,000 m^3/day extension to the Lakeview Water Pollution Control Plant (WPCP) located in Mississauga, Ontario.

The aeration tanks of the existing 224,000 m^3/day plant utilize sparjer coarse bubble diffusers. This form of aeration while having a low capital cost and low maintenance cost is relatively inefficient with an approximate 4% oxygen transfer efficiency.

Because of rising energy costs it was felt that the new extension should, if possible, utilize a more efficient aeration system than the existing coarse bubble sparjer diffusers. Fine bubble dome diffusers which offer an oxygen transfer efficiency of approximately 12% appeared an attractive option.

Gore & Storrie has designed and put into succesful operation two sewage treatment plants that utilize fine bubble dome aeration. These plants are located in Woodstock, Ontario and Kingston Township, Ontario. These plants have been operating successfully for 5 and 4 years respectively.

During the initial design period of the Lakeview extension it was established through literature search and technology transfer seminars that fine bubble aeration systems can slime in specific instances.

One current hypothesis is that high soluble biochemical oxygen demand (BOD) concentrations in aeration tank influent is a major contributor to slime growth on fine bubble diffusers.

Because of the high soluble BOD characteristics at the Lakeview plant it was decided that some pilot work with fine bubble diffusers would be prudent. Prior to finalizing the design of the fine bubble aeration system pilot studies were undertaken to determine whether fine bubble diffusers would slime when used to treat the sewage at the Lakeview W.P.C.P.

GLOSSARY

BOD - Biochemical oxygen demand
BOD_5-t - Total BOD, measured over an incubation period of 5 days at 20°C
BOD_5-s - soluble BOD, measured over an incubation period of 5 days at 20°C
$CBOD_5$-s - soluble carbonaceous BOD measured over an incubation period of 5 5 days at 20°C
D.O. - dissolved oxygen
FBDD - fine bubble dome diffuser,
HRT - Hydraulic retention time
HTL - heat treatment liquor
MLSS - mixed liquor suspended solids
OTE - oxygen transfer efficiency
OUR - oxygen uptake rate

PE - primary effluent
RAS - return activated sludge
WAS - waste activated sludge
WPCP - water pollution control plant.

DESCRIPTION OF EXISTING PLANT

The Lakeview WPCP consists of three sewage treatment units, plant 1, plant II and plant III. Each provide primary treatment followed by activated sludge secondary treatment.

The organic loading of plant II is typically higher than that of plant III for the following two reasons:

1. Recycle streams of heat treat liquor (HTL) from an on site Zimpro thermal conditioning unit and supernatant from the anaerobic digesters is recycled to plant II. The side streams have a soluble BOD of 3000 - 4000 mg/l.
2. The BOD of raw sewage influent to plant II is significantly higher than that of plant III because the plant inlet configuration channels the majority of the wastewater discharged from a local starch manufacturing facility to plant II. This industrial discharge is intermittent but does contribute significantly to the soluble BOD entering to plant II.

STAGES OF PILOT TESTING

For simplicity of description the pilot testing work can be divided into two stages. Stage 1 consisted of a series of eight reactor runs. The reactor runs commenced in June 1982 and ran to the fall of same year. Stage II of the pilot work ran from March to June 1983 and consisted of work carried out in conjunction with Napier-Reid Limited and Greey-Lightnin Ltd. both suppliers of fine bubble aeration devices.

Stage I

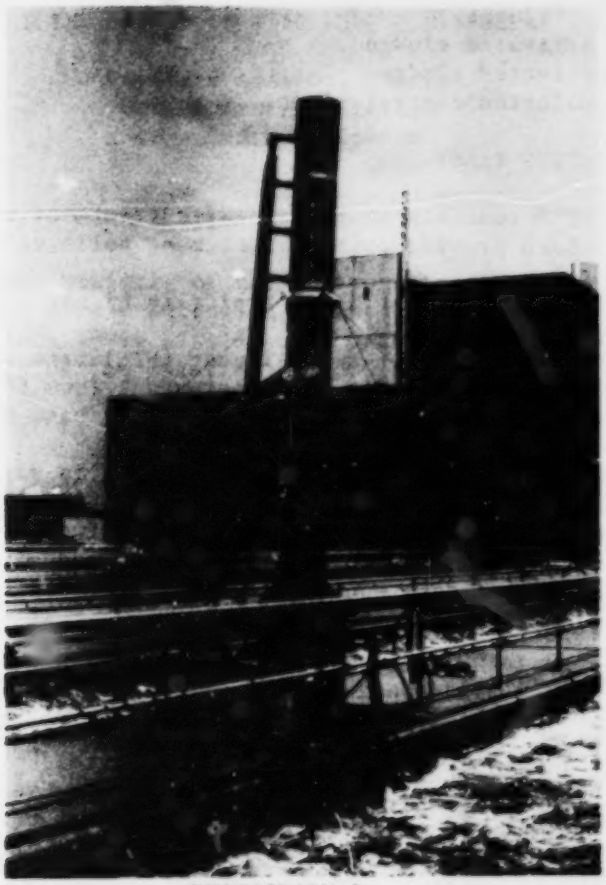
Eight reactor runs in stage I consisted of seven utilizing Norton/Gray type ceramic fine bubble domes and one utilizing a Nokia polyethylene fine bubble disc.

Study Equipment. Each reactor consisted of a fine bubble diffuser mounted at the bottom of a 350 mm diameter 4.3 metre high reactor.

Air was metered to the diffuser through a rotometer. Pressure drop across the diffuser was monitored with a manometer.

Metering pumps were used to pump process streams into the bottom of the reactor. Liquid overflowed from the top of the reactor.

Photograph No. 1 illustrates the study equipment.



PHOTOGRAPH 1
TEST REACTOR

Study Methods. Upon selection of operating criteria the reactors where placed in operation and monitored on a daily basis to ensure consistent operating conditions.

Samples of reactor feed streams and reactor contents were taken twice weekly. Dissolved oxygen, temperature, and oxygen uptake rates of the reactor contents were also monitored twice weekly.

Reactors were taken out of operation weekly and photographs taken of the diffusers.

Samples were analyzed in accordance with the 15th edition of Standard Methods.

Reactor 1. Reactor 1 equipped with a ceramic diffuser was operated under "worst case" conditions to determine if slime growth would occur on a diffuser at the inlet zone of plug flow aeration tank at Lakeview.

Mixed liquor from the inlet zone of a plant II aeration tank was pumped into the reactor at a rate of 200 litres per minute providing a hydraulic retention time (HRT) of 1.5 minutes.

The reactor feed soluble BOD varied significantly ranging from 43 mg/l to

204 mg/l, averaging 95 mg/l. Reactor content dissolved oxygen (DO) level averaged approximately 1 mg/l. Table 1 summarizes data obtained from reactor 1.

Slime growth was apparent in the second week of operation. Slime growth steadily increased reaching a thickness of 6-8 mm after four weeks. The surface of the slime growth was brownish-gray on the surface and black at the diffuser surface.

No significant slime growth was noted on the dome support saddle or reactor interior.

Photograph No. 2 illustrates a clean ceramic diffuser. Photograph No. 3 illustrates a slimed ceramic diffuser after 32 days of operation.

It was concluded that under "worst case" conditions, slime growth is rapid and significant.

Because of these results it was decided to model conditions more representative of the proposed complete mix aeration tank and retrofit of an existing plug flow aeration tank.

TABLE 1 - CERAMIC DOME DIFFUSER-REACTOR 1

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT		REACTOR FEED		REACTOR EFFLUENT				
			BOD_5-t	BOD_5-s	FLOW RATE	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s
	L/min	Kpa			L/min		°C			mg/L/hr	
1	25	-	-		228	-	-	-	-	-	-
7	28	-	-		209	47	19	1.5	1250	40	47
12	25	-	-		209	-	19	<0.5	1460	138	-
15	28	37.5	200	-	209	44	21	<0.5	2160	>300	31
19	30	37.2	-	-	205	150	20	0.8	2040	192	159
21	28	36.6	275	216	200	85	21	0.6	2710	150	92
26	26	-	218	213	205	42	21	0.8	2210	144	45
29	28	-	280	180	209	100	21	0.6	1990	120	130
33	20	37.5	221	125	182	90	23	2.3	1410	66	87
36	24	39	221	125	180	204	22	2.2	1150	59	195
AVG.	27	-	236	172	204	95	21	-	1820	134	98

Notes (1) All analytical results shown in mg/L.
(2) HRT 1.5 minutes

PHOTOGRAPH 2
CLEAN CERAMIC DIFFUSER

PHOTOGRAPH 3, REACTOR 2
SLIMED CERAMIC DIFFUSER AFTER 32 DAYS OF OPERATION

Reactor 2. Reactor 2 was equipped with a ceramic diffuser and operated to simulate the inlet conditions of the proposed complete mix aeration tank. A HRT of 60 minutes was selected with reactor feed consisting of primary effluent (PE)

and return activated sludge (RAS) from plant II. A RAS rate of 30% was selected.

The soluble BOD of the reactor contents was significantly lower than that of reactor 1 ranging from 13 mg/l to 68 mg/l, averaging 43 mg/l. Reactor content DO level averaged approximately 6 mg/l.

Table 2 summarizes data obtained from reactor 2.

Slime growth was evident after a week and remained to the end of the operating period. The thickness of the slime fluctuated but always remained significant.

Photograph 4 shows the diffuser from reactor 2 at the start. Photograph 5 shows the diffuser after seven days of operation.

TABLE 2 - CERAMIC DOME DIFFUSER-REACTOR 2

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT			RETURN ACT. SLUDGE		REACTOR EFFLUENT					
			FLOW RATE	BOD_5-t	BOD_5-s	FLOW RATE	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa	L/min			L/min		°C			mg/L/hr		
0	31	37.9	4.5	-	-	1.4	-	-	-	-	-	-	-
5	28	37.9	4.5	263	203	1.4	22	21	6.5	1520	34	62	-
7	28	37.9	4.5	235	206	1.4	30	22	7.8	1860	30	52	20
11	28	37.9	4.5	246	192	1.4	22	23	6.0	1050	40	40	-
14	28	37.9	4.5	341	318	1.4	28	22	4.8	1190	48	68	41
18	28	37.9	4.5	285	252	1.4	18	21	6.0	1320	72	25	12
21	-	-	-	224	200	-	-	-	-	-	-	-	-
25	28	37.9	4.5	160	101	1.4	18	21	6.2	1460	34	13	9
AVG.	28	-	4.5	251	210	1.4	23	21	-	1400	43	43	23 (54%)

Notes: (1) all analytical results shown in mg/L.
(2) HRT 60 minutes

PHOTOGRAPH 4, REACTOR 2
CERAMIC DIFFUSER AT BEGINNING OF TEST RUN

PHOTOGRAPH 5, REACTOR 2
DIFFUSER AFTER 7 DAYS OF OPERATION

Reactor 3. Reactor 3 equipped with a ceramic diffuser was operated concurrently with reactor 2 and was to simulate the inlet conditions expected with the retrofit of an existing plug flow aeration tank. A HRT of 20 minutes was selected. Mixed liquor from the outlet end of a plant II aeration tank was combined with PE.

The soluble BOD ranged from 19 mg/l to 58 mg/l, averaging 34 mg/l. The DO level of the reactor averaged approximately 6 mg/l.

Slime growth was significant after a week and was similar to that of reactor 2.

Table 3 summarizes data obtained from reactor 3.

Photograph 6 illustrates the diffuser of reactor 3 after 28 days of operation.

TABLE 3 - CERAMIC DOME DIFFUSER-REACTOR 3

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT			MIXED LIQUOR(3)			REACTOR EFFLUENT					
			FLOW RATE	BOD_5-t	BOD_5-s	FLOW RATE	D.O.	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa	L/min			L/min			°C			mg/L/hr		
4	35.2	-	4.5	277	176	13.2	7.0	26	21	6.0	1420	33	43	-
7	28.2	-	4.5	301	171	13.2	10.4	25	22	8.2	1090	30	53	-
11	28.2	-	4.5	166	134	13.2	9.3	27	22	8.0	1130	17	58	-
14	28.2	-	4.5	235	206	13.2	11.0	19	22	9.8	2840	24	43	21
18	21.1	37.9	4.5	246	192	13.2	8.2	21	23	7.2	1490	23	15	-
21	28.2	37.9	4.5	341	318	13.2	9.5	22	21	6.5	1510	30	54	-
25	28.2	37.9	4.5	285	252	13.2	6.5	34	21	6.0	920	34	28	18
28	28.2	37.9	4.5	224	200	13.2	6.0	32	21	7.5	1730	32	17	7
32	28.2	37.9	4.5	160	101	13.2	6.5	25	21	6.5	1300	25	13	4
35	28.2	37.9	4.5	-	-	13.2	5.0	34	23	4.5	-	-	-	-
36	-		4.5	186	156	13.2	-	11	-	-	1150	34	19	17
AVG.	28.2	-	4.5	242	191	13.2	-	25	22	-	1460	28	34	18 (54%)

Notes (1) All analytical results in mg/L
(2) HRT 20 minutes
(3) Mixed liquor was pumped from the aeration tank outlet

PHOTOGRAPH 6, REACTOR 3
CERAMIC DIFFUSER AFTER 28 DAYS OF OPERATION

Reactor 4. Reactor 4 was equipped with a polyethylene diffuser disc and was a repeat of "worst case" conditions noted for reactor 1.

Only minor slime growth was noted during the 42 days of operation. It was noted that the pressure drop across the diffuser increased significantly over time as noted in Table 4. This increase in head loss did not occur in the previous reactor runs equipped with ceramic diffusers. When the polyethylene diffuser was removed and examined a slurry of water and solids was noted to have accummulated on the underside of the diffuser. The diffuser had assumed a convex shape.

Photograph No. 7 illustrates the polyethylene disc after 14 days of operation.

TABLE 4 - POLYETHYLENE DISC DIFFUSER-REACTOR 4

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT		REACTOR FEED		REACTOR EFFLUENT					
			BOD_5-t	BOD_5-s	FLOW RATE	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa			L/min		°C			mg/L/hr		
4	64.8	36.4	131	59	190	29	21	6.0	1570	44	25	
7	62.0	36.4	-	-	190	-	21	4.0	-	-	-	
11	50.7	36.9	174	137	190	89	20	3.8	1360	66	24	
14	59.2	36.9	247	171	190	38	21	5.8	1190	26	61	
19	64.8	36.4	263	203	190	28	22	6.5	4440	46	63	
21	59.2	36.6	235	194	190	34	22	7.2	5400	20	29	
25	56.4	36.6	261	182	190	24	23	6.2	4240	37	31	
26	56.4	39.6	-	-	190	-	-	-	-	-	-	
28	56.4	40.8	334	300	190	21	22	6.5	3550	32	14	6
32	42.3	41.8	165	159	190	45	21	3.5	1540	85	44	25
35	53.6	41.3	203	192	190	56	21	3.5	1690	102	36	30
39	42.3	41.8	190	126	190	32	20	3.5	2030	80	30	24
42	53.6	41.8	214	183	190	47	21	2.3	2340	114	46	34
	55.5	-	220	173	190	40	21	-	2670	59	37	24 (65%)

Notes (1) All analytical results shown in mg/L
(2) HRT 1.5 minutes
(3) This reactor was equipped with a NOKIA diffuser

PHOTOGRAPH 7
POLYETHYLENE DISC DIFFUSER AFTER 14 DAYS OF OPERATION

Interim Discussion. With the completion of reactor runs 1-4 it became apparent that fine bubble diffusers had a high propensity to slime at Lakeview. It was decided to run another series of tests under similar conditions to those noted for reactors 2 and 3 except that recycle HTL would be eliminated from plant influent. It was hoped that the elimination of heat treat liquor would reduce the soluble BOD entering the aeration tanks and therefore eliminate slime growth on fine bubble diffusers.

This elimination of HTL was accomplished by moving the reactors to plant III which receives no HTL. A third reactor was set up as a control which did receive a side stream of HTL.

Reactors 5, 6. These reactors were equipped with ceramic diffusers operated under the same conditions as noted for reactors 2 and 3 except with the absence of HTL.

The diffusers of both reactors slimed significantly over the 5 weeks of operation.

Table 5 and 6 summarize data obtained from reactor 5 and 6.

TABLE 5 - CERAMIC DOME DIFFUSER-REACTOR 5

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT			RETURN ACT. SLUDGE		REACTOR EFFLUENT					
			FLOW RATE	BOD_5-t	BOD_5-s	FLOW RATE	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa	L/min			L/min		°C			mg/L/hr		
1	14.2	34.2	12.3	-	-	5.0	-	-	-	-	-	-	-
6	14.2	34.2	12.3	114	106	5.0	14	20	<0.5	1490	-	35	32
8	14.2	-	12.3	262	216	5.0	11	18.5	<0.5	1720	74.2	99	72
12	14.2	34.2	12.3	126	117	5.0	<10	18.5	7.0	2320	24.0	<10	<5
15	14.2	34.2	12.3	218	189	5.0	12	18	3.7	2600	30.0	15	8
19	14.2	34.2	12.3	186	180	5.0	11	18	2.5	2380	42.0	19	15
22	14.2	34.9	12.3	214	159	5.0	34	16.5	4.5	2210	-	40	13
23	-	-	-	-	-	5.0	-	-	-	-	24.0	-	-
26	14.2	34.2	12.3	152	102	5.0	6	19	5.5	2100	25.2	17	6
29	14.2	34.2	12.3	290	246	5.0	20	17	5.7	2370	43.2	54	26
33	14.2	34.2	12.3	230	210	5.0	5	16	3.5	1840	28.8	40	33
36	14.2	34.2	12.3	258	228	5.0	6	18	2.6	1650	33.6	60	52
AVG.	14.2	-	12.3	205	175	5.0	13	18	-	2070	35	39	25 (65%)

Notes (1) All analytical results shown in mg/L.
(2) HRT 20 minutes

TABLE 6 - CERAMIC DOME DIFFUSER-REACTOR 6

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT			RETURN ACT. SLUDGE		REACTOR EFFLUENT					
			FLOW RATE	BOD_5-t	BOD_5-s	FLOW RATE	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa	L/min			L/min		°C			mg/L/hr		
1	14.2	34.2	4.1	-	-	1.9	-	-	-	-	-	-	-
6	14.2	34.2	4.1	114	106	<1.9	14	20	<0.5	685	-	63	60
8	14.2	34.2	4.1	262	216	<1.9	11	18.5	4.5	1970	28.2	13	12
12	14.2	34.2	4.1	126	117	1.9	<10	18.5	7.0	2140	31.2	<10	<5
15	14.2	34.2	4.1	218	189	1.9	12	18	4.5	1820	21.6	19	15
19	14.2	34.2	4.1	186	180	<1.9	11	18	2.7	1130	21.6	30	20
22	14.2	34.2	4.1	214	159	1.9	34	16	6.5	1050	-	48	14
23	-	-	-	-	-	-	-	-	-	-	12.0	-	-
26	14.2	34.2	4.1	152	102	<1.9	6	17.5	4.5	930	18.0	38	28
29	14.2	34.2	<4.1	290	246	<1.9	20	16	6.2	1580	25.2	46	34
33	14.2	34.2	4.1	230	210	1.9	5	14.5	5.5	970	24.0	67	60
36	14.2	34.2	4.1	258	228	1.9	6	17	5.5	1380	25.2	34	5
AVG.	14.2	34.2	4.1	205	175	1.9	13	17.4	-	1370	23	37	25 (68%)

Notes (1) All analytical results shown in mg/L.
(2) HRT 60 minutes

Reactor 7. Reactor 7 was operated concurrently with reactors 5 and 6. Conditions of operation where similar to reactor 5 except that HTL was added. Slime growth on the diffuser was distinct after the 5 weeks of operation.

Table 7 summarizes data obtained from reactor 7.

TABLE 7 - CERAMIC DOME DIFFUSER-REACTOR 7

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT(3)			RETURN ACT. SLUDGE		REACTOR EFFLUENT					
			FLOW RATE	BOD_5-t	BOD_5-s	FLOW RATE	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa	L/min			L/min		°C			mg/L/hr		
1	14.2	34.2	12.3	-	-	5.0	-	20	-	-	-	-	-
6	17.3	34.2	12.3	251	195	5.0	15	20	<0.5	2220	46.2	65	62
8	17.3	34.2	12.3	191	177	5.0	8	18.5	1.0	2270	40.2	36	30
12	17.3	34.9	12.3	212	203	5.0	<10	18.5	5.0	1610	25.2	<10	6
15	17.3	34.9	12.3	206	180	<5.0	<10	19	3.0	1400	79.8	68	60
19	26.4	34.9	12.3	150	78	5.0	15	17.5	4.5	1670	34.2	30	26
22	22.2	34.9	12.3	175	119	5.0	25	16	6.0	1920	22.8	38	13
26	22.2	34.2	12.3	253	221	<5.0	6	17.5	5.0	1100	36.0	168	144
29	22.2	34.2	12.3	196	192	<5.0	11	17	4.5	2480	54.0	54	22
33	42.6	34.2	12.3	214	205	<5.0	<5	16	5.0	2070	28.8	105	99
AVG.	-	-	12.3	205	174	5.0	12	18	-	1860	40.8	64	50 (75%)

Notes (1) All analytical results shown in mg/L
(2) HRT 20 minutes
(3) This reactor also received 0.14 L/min of HTL @ BOD_5 of 4000 mg/L; the total feed characteristics were: BOD_5-t = 248 mg/L and BOD_5-s = 217 mg/L

Interim Discussion. To determine if any feed condition would result in no slime growth it was decided to feed aeration tank effluent to a reactor.

Reactor 8. Reactor 8 was operated with a feed of aeration tank effluent with a hydraulic retention time of 1.5 minutes. Soluble BOD in reactor effluent ranged from 5 mg/l to 30 mg/l, averaging 12 mg/l. No slime growth was noted after 42 days of operation. The diffuser remained clean and gritty to the touch.

Table 8 summarizes data obtained from reactor 8.

TABLE 8 - CERAMIC DOME DIFFUSER-REACTOR 8

STUDY DAY	AIR FLOW RATE	AIR PRESS	PRIMARY EFFLUENT		REACTOR FEED		REACTOR EFFLUENT					
			BOD_5-t	BOD_5-s	FLOW	BOD_5-s	TEMP.	D.O.	MLSS	OUR	BOD_5-s	$CBOD_5$-s
	L/min	Kpa			L/min		°C			mg/L/hr		
2	17.4	-	-	-	230	10	15.0	10.0	2240	16	11	9
7	17.4	-	-	-	230	9	16.0	8.2	2910	19	9	5
9	17.4	-	-	-	230	6	16.0	8.2	2380	16	7	6
15	23.1	-	110	63	260	17	15.0	8.9	2100	14	13	12
22	18.7	-	55	30	260	5	13.0	9.5	1580	5	5	5
24	18.7	-	111	39	260	11	13.5	9.7	2000	10	13	11
29	18.7	-	-	-	260	11	15.0	9.8	1980	12	12	12
31	18.7	-	150	120	260	15	15.0	9.2	1980	14	17	12
35	18.7	-	-	-	260	30	15.0	8.0	2290	28	30	25
38	18.7	-	199	162	260	7	14.5	8.5	1860	18	7	7
42	18.7	-	-	-	260	8	13.0	7.8	2010	17	8	4
AVG.	18.7	-	125	83	252	12	14.6	-	2121	15	12	10

Notes (1) All analytical results shown in mg/L

Oxygen Transfer Efficiency. After establishing that diffusers had a high propensity to slime in Lakeview sewage it was decided to determine if the oxygen transfer efficiency of diffusers is affected by the slime growth.

Manufacturers literature indicates that an oxygen transfer efficiency of 25% could be expected with a clean diffuser in tap water. Testing of a clean diffuser averaged 25.8%.

A lightly slimed diffuser was tested. The oxygen transfer efficiency in tap water averaged 24.1%.

A heavily slimed dome was tested. The oxygen transfer efficiency in tap water averaged 22.7%. This represents a 12% reduction in efficiency over a clean diffuser.

This relatively small loss of efficiency is surprising and may indicate that sliming may not be as detrimental as first assumed, however the tests were not of long enough duration for a firm conclusion.

Table 9 illustrates oxygen transfer efficiency results obtained.

TABLE 9 - FBDD STUDY OXYGEN TRANSFER TEST RESULTS ON CLEAN AND SLIME-COATED CERAMIC DOMES

DATE OF TEST (1982)	DESCRIPTION OF FBDD TESTED		OXYGEN TRANSFER EFFICIENCY (%)
Nov. 23	New, clean dome	run 1 run 2 run 3	25.1 25.6 26.6
		Avg.	25.8
Nov. 23	Lightly slimed dome	run 1 run 2	22.8 25.8
		Avg.	24.1
Dec. 3	Heavily slimed dome	run 1 run 2	22.5 22.9
		Avg.	22.7

Interim Discussion. The conclusion of the oxygen transfer efficiency tests concluded stage I of the pilot testing work.

It was apparent from Stage I of the pilot testing program that the causes and effects of slime growth on fine bubble diffusers had to be much better defined before fine bubble diffusers would be desirable in a plant with a sewage having the high soluble BOD characteristics found at Lakeview.

Stage II

To provide some energy savings over coarse bubble diffusers it was decided to test if fine bubble diffuser tubes running at relatively high air flow rates would create enough turbulence to prevent slime growth from adhering to the diffuser material. A Pearlcombe fine bubble tube diffuser as distributed by Napier-Reid was chosen for initial review. Literature indicates these diffusers running at high air flow rates afforded a potential oxygen transfer efficiency in sewage of 8% with moderate increase in capital cost over coarse bubble sparjer diffusers. This compares to a 4% efficiency normally exhibited by coarse bubble diffusers and represents a significant potential energy savings. A proposal was solicited from Napier-Reid to test their Pearlcombe tube diffuser at Lakeview prior to consideration for the secondary treatment extension.

Napier-Reid agreed to test their Pearlcombe diffuser at Lakeview and also requested permission to test two other similar diffuser tubes concurrently. In consultation with the Ministry it was agreed that Napier Reid would set up and maintain the test equipment and that the Ministry would fund the laboratory testing work carried out by Gore & Storrie Ltd.

Greey-Lightnin who markets the Nokia fine bubble tube diffuser, a unit similar to the Pearlcombe, was contacted and informed of an opportunity to pilot test tube diffusers at Lakeview.

Greey-Lightnin in a meeting with the Ministry and Gore & Storrie Ltd. outlined a pilot test program of running diffuser tubes and discs with a view to controlling slime growth by the application of formic acid. This technique has had apparent success in Europe against inorganic fouling and it was Greey-Lightnin's belief the technique would be effective against sliming at Lakeview.

The Ministry agreed to permit the piloting of tube and disc diffusers at Lakeview. Test reactors constructed for stage I of the diffuser pilot work were utilized for the disc testing.

Napier-Reid Pilot Work. The Napier-Reid pilot equipment was setup at Plant III. Plant III at the time of this work ran as a base-loaded conventional plant with primary effluent and return sludge introduced at the head of pass 1.

The testing equipment consisted of three diffusers mounted approximately 2 metres below the aeration tank surface and fed metered air through individual 20 mm pipes. Photograph No. 8 illustrates the testing equipment complete with diffusers after seven days of operation.

PHOTOGRAPH 8
NAPIER-REID DIFFUSERS AFTER 7 DAYS OF OPERATION

The diffusers tested were the Pearlcombe SP35 (9-1), Napier-Reid HUT 600 (9-2) and MIZPE (9-3).

The Napier-Reid diffusers were initially installed at the end of pass 1 the quarter point of Plant III aeration tank. Within two weeks sliming of tubes was evident. The diffusers were moved to the head of pass 1 to be more directly comparative to the Nokia testing program underway at Plant II. At the end of four weeks all three diffusers had extensive slime growth. No pressure increase was observed across any of the diffusers over the test period.

Table 10 and 11 summarizes data obtained on Napier-Reid Diffusers.

TABLE 10
NAPIER REID TUBE DIFFUSERS
LOCATED: PLANT 3 END OF PASS 1

		PRIMARY EFFLUENT					AERATION TANK MIXED LIQUOR						DIFFUSER 9-1 (PEARLCOMBE)		DIFFUSER 9-2 (HUT-600)		DIFFUSER 9-3 (MIZPE)	
Date 1983	Days	pH	S.S.	BOD_5-t	BOD_5-S	$CBOD_5$-S	Temp. °C	D.O.	O.U.R mg/L/hr	MLSS	BOD_5-S	$CBOD_5$-S	Air Flow Rate l/min	Air Press. kPa	Air Flow Rate l/min	Air Press. kPa	Air Flow Rate l/min	Air Press. kPa
03.18	0	6.9	116	223	173	117	14.5	1.3	37.8	3030	20	11	186.9	4.4	189.7	7.4	39.6	10.8
03.22	4	7.3	108	217	179	116	13.0	3.5	19.2	2480	26	9	186.9	N.A.	189.7	N.A.	39.6	N.A.
03.25	7	6.8	105	215	180	120	13.0	2.0	20.4	2080	20	11	186.9	N.A.	189.7	N.A.	39.6	N.A.
AVG		-	110	218	177	118	13.5	2.3	25.8	2530	22	10	186.9	-	189.7	-	39.6	-

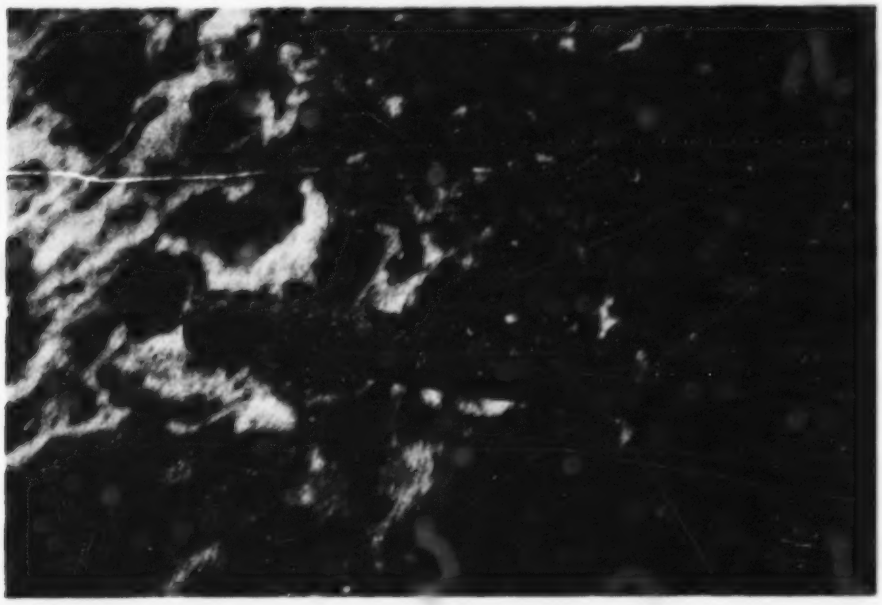
Notes: (1) All analytical results in mg/l
(2) Air pressures noted exclude static water pressure

TABLE 11
NAPIER REID TUBE DIFFUSERS
LOCATED: PLANT 3 START OF PASS 1

		PRIMARY EFFLUENT					AERATION TANK MIXED LIQUOR						DIFFUSER 9-1 (PEARLCOMBE)		DIFFUSER 9-2 (HUT 600)		DIFFUSER 9-3 (MIZPE)	
Date 1983	Days	pH	S.S.	BOD_5-t	BOD_5-S	$CBOD_5$-S	Temp. °C	D.O.	O.U.R mg/l/hr	MLSS	BOD_5-S	$CBOD_5$-S	Air Flow Rate l/min	Air Press. kPa	Air Flow Rate l/min	Air Press. kPa	Air Flow Rate l/min	Air Press. kPa
03.29	11	6.6	95	204	170	141	13.0	2.5	90	2280	72	N.A.	206.7	N.A.	212.3	N.A.	53.8	N.A.
03.31	13	6.8	105	215	185	175	13.0	2.3	90	2310	180	N.A.	189.7	3.9	178.4	7.6	27.5	9.8
04.05	18	6.9	108	230	182	150	14.0	2.3	90	2400	102	90	195.4	2.9	203.9	6.1	50.9	7.1
04.08	21	7.2	72	179	169	146	14.0	3.2	78	2290	74	66	195.4	3.4	206.7	7.6	50.9	7.1
04.12	25	7.2	102	213	188	176	14.0	3.3	76	1950	93	74	195.4	3.7	206.7	7.4	50.9	7.1
04.15	28	6.9	120	144	107	101	14.0	3.0	50	2400	41	39	195.4	2.9	203.9	4.9	50.9	7.1
04.19	32	7.1	221	156	99	78	14.0	2.8	45	2676	47	38	195.4	3.4	206.7	7.6	50.9	5.6
04.22	35	7.1	130	239	164	138	15.0	1.3	102	1232	72	60	195.4	2.9	206.7	7.1	50.9	4.9
04.26	39	6.8	65	279	219	206	15.5	1.0	84	2528	77	74	195.4	2.9	206.7	7.6	50.9	5.6
04.29	42	6.6	156	295	216	204	16.0	0.75	200	2490	90	87	195.4	2.9	206.7	8.3	50.9	4.4
05.03	46	7.2	272	214	92	77	15.0	1.50	42	1592	24	18	195.4	2.9	206.9	7.8	50.9	4.9
AVG		-	131	215	163	144	14.0	2.2	86	2201	76	61	195.4	-	7.2	- -	1.7	-

Notes: (1) All analytical results in mg/l
(2) Air pressures noted exclude static water pressure

Photograph No. 9 illustrates slimed diffuser tubes after 28 days of operation.



PHOTOGRAPH 9
NAPIER-REID DIFFUSERS AFTER 28 DAYS OF OPERATION

The test rig was moved to the head of pass 3, the halfway point of the aeration tank, and clean HUT 600 (12-2) and MIZPE (12-3) diffusers installed. Normal monitoring work was continued. After four weeks both diffusers were observed to have small patches of slime. The sliming appeared to have stabilized. This indicates that some sliming can be expected to at least the mid-point of the tank.

A surprising observation was that while the diffusers did not slime significantly, a significant increase in pressure loss was observed over time across the diffusers. This increase in pressure drop across the diffusers had not previously been observed except in reactor run 4 of stage I. In the case of reactor 4 the increased headloss appeared to have been caused by air side fouling with mixed liquor entering the saddle assembly. The interior of the Naper-Reid diffusers appeared clean. No mixed liquor appeared to have found its way to the air side of the diffuser. It is suspected that the increase in headloss of the diffusers is related to iron deposits. Ferrous and Ferric Chlorides are added at the head of pass 3, the location of this test run. It is believed that soluble ferrous iron penetrates the diffuser pores and is oxidized causing a deposit of ferric compounds.

Table 12 summarizes the results obtained and illustrates the increase in headloss across the diffusers.

TABLE 12
NAPIER-REID TUBE DIFFUSER
DIFFUSERS LOCATED: PLANT 3, START OF PASS III

		PRIMARY EFFLUENT					AERATION TANK MIXED LIQUOR						DIFFUSER 12-2 (HUT 600)		DIFFUSER 12-3 (MIZPE)	
Date 1983	Days	pH	S.S	BOD_5-t	BOD_5-S	$CBOD_5$-S	Temp.	D.O.	OUR	MLSS	BOD_5-S	$CBOD_5$-S	Air Flow l/min	Air Press. kPa	Air Flow l/min	Air Press. kPa
05.10	0	7.2	176	260	180	177	15.5	3.5	42	2840	30	24	108	1.7	31.1	4.9
05.13	3	6.7	226	371	248	237	16.0	0.75	66	2950	46	36	110	4.4	31.1	4.9
05.17	7	-	-	-	-	-	N.A.	N.A.	-	-	-	-	110	4.9	31.1	12.0
05.20	10	6.7	120	174	134	120	16.0	2.6	28	2980	22	10	110	8.3	31.1	13.0
05.27	17	7.1	104	190	150	120	16.0	3.75	34	2630	13	4	113	10.2	31.1	14.2
06.03	24	6.6	201	275	210	120	17.0	2.25	30	2470	10	4	113	13.2	31.1	20.4
AVG		-	149	254	184	171	16.1	2.6	40	2774	24	16	111		31.6	

Note: (1) All analytical results in mg/l
(2) Air pressures noted exclude static water pressure

Observations of Slimed Tubes in Test Tank Filled with Tap Water. Napier-Reid transported the slimed diffusers to their clean water test tank.

A slimed pearlcombe diffuser was mounted in the test tank and the bubble pattern observed through the clear tank side. The bubbles were distinct and exited from discrete points over the entire surface of the tube.

A clean pearlcombe diffuser was mounted in the test tank and the bubble pattern observed. The air leaving the tube appeared as a cloud exiting the diffuser with no distinct bubbles. All air was exiting from the top surface of the tube.

It was readily apparent that the slimed pearlcombe tube produced larger bubbles with a presumed corresponding decrease in oxygen transfer efficiency.

Nokia Pilot Work. The Nokia pilot equipment was set-up on one of the four aeration tanks at Plant II. Plant II at the time of testing ran with the first pass acting as a return sludge reaeration zone with primary effluent introduced at the head of pass II. The pilot test equipment was set-up at the head of pass II.

Three fine bubble tube diffusers were installed in the aeration tank approximately 3 metres below liquid level in a similar manner to the Napier-Reid diffusers described previously. One diffuser was a control, one was run at low air flow rate to promote sliming and the remaining diffuser was to be acid cleaned.

All three tube diffusers rapidly developed slime growth; within two weeks

spotty sliming was observed. The slime on the control unit continued to accummulate throughout the six-week study period. No headloss increase was noted.

Formic acid injection into the air stream to the test unit (10-2) was started after slime growth was observed. The initial dosage of 15 grams of acid at day 14 seemed to have no appreciable effect. Subsequent doses of 30 grams of formic acid were used; initially as one large injection and on the last three occasions in small amounts spread over a 30-minute period.

During the last two weeks a major reduction in slime growth was achieved, only a very thin coating of slime remained on the test unit while the sliming on the control unit continued unabated.

Table 13 summarizes data obtained for the Nokia tube diffusers.

TABLE 13
GREEY-LIGHTNIN TUBE DIFFUSERS
LOCATED: PLANT 2, AERATION TANK 6, HEAD OF PASS II

		PRIMARY EFFLUENT					AERATION TANK MIXED LIQUOR						10-1 (NOKIA)		10-2 (NOKIA)			10-3 (NOKIA)		
Date 1973	Days	pH	S.S.	BOD_5-t	BOD_5-S	$CBOD_5$-S	Temp. °C	D.O.	O.U.R. mg/l/hr	MLSS	BOD_5-S	$CBOD_5$-S	Air Flow l/min	Air Press. kPa	Air Flow l/min	Air Press. kPa	Acid Dosage m/s	Air Flow l/min	Air Press. kPa	Acid Dosage m/s
04.22	0	6.8	96	164	120	98	14.0	1.75	-	1660	45	43	N.A.	7.8	N.A.	8.0	-	N.A.	8.1	-
04.26	4	-	-	-	-	-	-	-	-	-	-	-	N.A.	10.8	N.A.	10.8	-	N.A.	10.3	-
04.29	7	7.0	130	233	177	171	16.0	0.60	125	1780	93	90	223.7	6.6	223.7	6.6	-	184.1	7.1	-
05.03	11	-	-	-	-	-	-	-	-	-	-	-	206.7	7.8	198.2	7.8	-	164.2	7.4	-
05.06	14	-	-	-	-	-	-	-	-	-	-	-	257.7	7.4	260.5	8.3	15	99.1	6.1	-
05.10	18	-	-	-	-	-	-	-	-	-	-	-	257.7	7.4	257.7	6.4	-	99.1	6.1	-
05.13	21	6.8	146	269	227	197	16.5	0.50	130	1950	93	88	257.7	6.6	257.7	6.4	30	99.1	6.1	-
05.17	25	-	-	-	-	-	-	-	-	-	-	-	260.5	3.2	260.5	3.2	-	99.1	3.0	-
05.20	28	-	-	-	-	-	-	-	-	-	-	-	223.7	3.2	260.5	2.7	30	99.1	4.5	-
AVG	4 wks	-	124	222	175	155	15.5	0.95	128	1797	77	74	241.1	-	245.5	-	-	120.5	-	-
05.24	32	-	-	-	-	-	-	-	-	-	-	-	237.9	3.5	223.7	3.2	-	96.3	3.2	-
05.27	35	-	-	-	-	-	-	-	-	-	-	-	257.7	3.0	254.9	4.5	30	96.3	2.2	-
06.01	40	-	-	-	-	-	-	-	-	-	-	-	237.9	3.0	229.4	2.2	-	99.1	7.9	-
06.03	42	-	-	-	-	-	-	-	-	-	-	-	260.5	3.2	260.5	3.2	30	96.3	2.0	-
AVG	6 wks	-	124	222	175	155	15.5	0.95	128	1797	77	74	243.8	-	244.3	-	-	112.0	-	-

Notes: (1) All analytical results in mg/l. (2) Air pressures noted exclusive static water pressure

Concurrent with the above tube diffuser tests, two diffuser discs were run in the same reactors that were used for stage I. Mixed liquor from the head of pass II was pumped into the reactors. Aeration tank soluble BOD averaged 77 mg/l. After several weeks of operation, small amounts of sliming were noted on both diffusers. However, the amount of slime on the diffusers did not increase significantly after six weeks of operation.

Table 14 summarizes data obtained for the Nokia disc diffuser.

TABLE 14
GREEY-LIGHTNIN DISC DIFFUSERS
LOCATED: PLANT 2, AERATION TANK 6, HEAD OF PASS II

		PRIMARY EFFLUENT					AERATION TANK MIXED LIQUOR						11-1 (NOKIA)		11-2 (NOKIA)		
Date 1983	Days	pH	S.S.	BOD_5-t	BOD_5-S	$CBOD_5$-S	Temp. °C	D.O.	O.U.R. mg/l/hr	MLSS	BOD_5-S	$CBOD_5$-S	Air Flow l/min	Air Press. kPa	Air Flow l/min	Air Press. kPa	Acid Dosage m/s
04.22	0	6.8	96	164	120	98	14.0	1.75	-	1660	45	43	73.6	4.9	65.1	5.4	-
04.26	4	-	-	-	-	-	-	-	-	-	-	-	73.6	5.1	67.9	5.9	-
04.29	7	7.0	130	233	177	171	16.0	0.60	125	1780	93	90	73.6	3.7	65.1	4.9	-
05.03	11	-	-	-	-	-	-	-	-	-	-	-	73.6	8.1	56.6	11.2	-
05.06	14	-	-	-	-	-	-	-	-	-	-	-	56.6	7.8	56.6	7.1	5
05.10	18	-	-	-	-	-	-	-	-	-	-	-	59.4	5.1	56.6	10.0	-
05.13	21	6.8	146	269	227	197	16.5	0.50	130	1950	93	88	59.4	6.1	56.6	11.5	10
05.17	25	-	-	-	-	-	-	-	-	-	-	-	56.6	2.2	56.6	2.2	-
05.20	28	-	-	-	-	-	-	-	-	-	-	-	56.6	2.2	56.6	2.2	10
AVG	4 wks	-	124	222	175	155	15.5	0.95	128	1797	77	74	64.7	-	59.7	-	-
05.24	32	-	-	-	-	-	-	-	-	-	-	-	56.6	4.2	56.6	2.7	-
05.27	35	-	-	-	-	-	-	-	-	-	-	-	56.6	2.7	56.6	2.5	10
06.01	40	-	-	-	-	-	-	-	-	-	-	-	56.6	1.7	56.6	2.5	-
06.03	42	-	-	-	-	-	-	-	-	-	-	-	56.6	2.5	56.6	2.5	10
AVG	6 wks	-	124	222	175	155	15.5	0.95	128	1797	77	74	62.3	-	58.8	-	-

Notes: (1) All analytical results in mg/l, (2) Air pressures noted exclude static water pressure

Formic acid was injected on five occasions into the air stream of the test unit. Any improvement due to this acid injection could not be documented because no significant slime growth was evident on either diffuser. It is of interest to note that similar amounts of sliming were observed during tests with a Nokia disc diffuser during reactor 4 run of stage I.

CONCLUSIONS

As a result of the testing done at the Lakeview Water Pollution Control Plant the following conclusions can be drawn.

1. At Lakeview sliming can be expected on fine bubble diffusers when installed in the inlet zone of the plug flow aeration tank.
2. Sliming is not expected at the end of a plug flow aeration tank as long as soluble BOD_5 remains below approximately 15 mg/l.
3. Sliming can be expected to approximately the halfway point of an aeration tank at the Lakeview WPCP
4. There is no evidence to suggest that fine bubble tube diffusers will be less likely to slime than fine bubble domes or discs.

5. It appears that sliming will occur on fine bubble diffusers when soluble BOD_5 levels exceed approximately 20 mg/l.
6. Sliming itself does not cause significant increase in pressure drop across fine bubble diffusers.
7. Headloss increases that do occur appear to be from air side fouling or inorganic compound build-up rather than slime growth.
8. The observed size of bubbles from slimed diffusers is significantly larger than that seen from a clean diffuser with a presumed decrease in oxygen transfer efficiency.
9. Formic acid cleaning has an appreciable effect on the extent of a diffuser sliming. Formic acid cleaning was observed to lessen but not eliminate sliming of diffusers.
10. It appears that ferrous chloride can precipitate out onto diffusers and cause significant increases in headloss across the diffuser.

RECOMMENDATIONS

1. Fine bubble diffusers not be installed in any plant with high soluble BOD influent without pilot testing.
2. The mechanism and the control of slime growth be investigated so that methods of avoiding or controlling slime growth can be developed.
3. The limits of soluble BOD under which diffusers can be operated without problems must be explored. When these limits have been identified, satisfactory operation must be demonstrated on a pilot scale to confirm that factors other than soluble BOD concentrations have no significance with respect to sliming. Tests must therefore be conducted to yield information on the relationship of the following factors to sliming.
   A. Soluble BOD concentration
   B. Aeration tank type, plug flow, complete mix, step aeration
   C. Activated sludge characteristics
   D. Air flow per diffuser
   E. Mixed liquor suspended solids concentration
   F. Dissolved oxygen concentration in mixed liquor

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DISPOSITION CONJOINTE DES BOUES RÉSIDUAIRES ET DES ORDURES MÉNAGÈRES PAR INCINÉRATION

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INTRODUCTION

Dans le cadre d'une étude préparatoire pour le traitement des eaux usées du secteur Est de la Communauté Urbaine de Québec[1], nous avons investigué la possibilité de disposer des boues résiduaires du traitement, à l'incinérateur à déchets solides existant de la C.U.Q.

L'étude a porté principalement sur les principes d'opération d'un tel système en vue d'en établir la faisabilité. Le calcul des bilans thermiques de même qu'une visite en Europe de deux (2) installations existantes (Wilrijk en Belgique et Lisieux en France) nous a permis de conclure en sa faisabilité.

Aussi, nous verrons ci-après, de façon simple et sans prétention, le cheminement effectué pour en arriver à conclure en faveur de cette option. Et considérant que ce mode de disposition des boues de traitement d'eaux usées s'avère intéressant pour la C.U.Q., ne pourrait-il pas en être ainsi à certains autres endroits au pays?

SOMMAIRE DU TRAITEMENT ÉTUDIÉ

Territoire et population desservie

L'usine d'épuration du secteur Est de la Communauté Urbaine de Québec desservira la majeure partie des villes de Québec et Charlesbourg, et la totalité des villes de Beauport et Vanier. Selon nos prévisions démographiques, ce territoire regroupera, en 2006, une population totale équivalente de 353 575 personnes (résidences, commerces et industries) et, actuellement, toutes les infrastructures sanitaires (collecteurs, diffuseurs, station de pompage) y sont existantes.

Caractéristiques des eaux usées

En raison d'un accroissement assez lent de la population à desservir et des corrections qui seront apportées aux réseaux d'égout, dans le futur, le débit moyen total acheminé à l'usine, en temps sec, sera sensiblement constant au cours des années, et égal à 2,1 mcs (1,38 mcs la nuit et 3,19 mcs en pointe horaire). En périodes pluvieuses, le débit maximal peut atteindre la capacité des postes de pompage Limoilou et St-Pascal, soit 7,62 mcs. D'un point de vue qualitatif, ces eaux usées sont caractérisées par des concentrations de l'ordre de 120 mg/l en matières organiques, de 133 mg/l en solides en suspension et de 3,6 mg/l en phosphates totaux.

Type de traitement

Sur la base de ces données et pour répondre aux seuls objectifs de cette étude préparatoire, une conception d'usine de traitement primaire physico-chimique a été réalisée pour atteindre les rendements d'épuration suivants: réductions de 60% des solides en suspension, de 50% des matières organiques, de 80% des phosphates totaux, de 99% des matières flottantes (huiles) et de

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1. Une deuxième usine d'épuration est prévue pour le secteur Ouest.

99,9% des bactéries pathogènes (été seulement).

Description de l'usine

L'usine d'épuration sera composée, tout d'abord, d'un prétraitement (dégrillage fin, dessablage, déshuilage) où toutes les eaux usées du territoire desservi seront admises (débit de 1,38 à 7,62 mcs). Ensuite, il y aura, en périodes de pluie, dérivation vers l'effluent, des débits supérieurs à la pointe horaire de temps sec (3,19 mcs), et traitement physico-chimique des eaux usées comprenant les étapes suivantes: mélange rapide du coagulant (75 mg/l d'alun), floculation, décantation et désinfection. Les eaux ainsi traitées seront finalement rejetées au fleuve St-Laurent par l'entremise d'un bassin d'effluent et du système de diffusion actuel. En fonction des débits véhiculés et des conditions de marées, les eaux d'effluent de l'usine seront déversées par gravité ou par pompage. Alors que les écumes et matières flottantes enlevées au niveau des dessableurs et des décanteurs seront épaissies par flottation et disposées par enfouissement sanitaire, les boues résultantes du traitement, soit 24,4 tonnes par jour de matières sèches, seront disposées comme ci-après.

Disposition des boues

La disposition des boues résiduaires d'une usine d'épuration peut être assurée par diverses méthodes, dont l'application est influencée par le type de boues et par les conditions propres à chaque installation. Jusqu'à maintenant, les alternatives les plus souvent retenues sont les suivantes: l'incinération ou la digestion avec disposition finale par enfouissement sanitaire ou amendement au sol (compostage). Pour chacune de ces méthodes, les applications sont diverses, dépendant des besoins et des conditions locales. Par exemple, dans certaines régions du sud des Etats-Unis, le méthane produit par la digestion des boues est vendu pour utilisation domestique. En Europe, et récemment aux Etats-Unis, des incinérateurs à déchets solides assurent la disposition combinée des ordures ménagères et des boues produites par des stations d'épuration urbaines.

Etant donné la proximité relative de l'incinérateur à déchets solides de la C.U.Q., par rapport au point de concentration des eaux usées du secteur Est (système de diffusion), soit environ 2 kilomètres, et compte-tenu d'une utilisation actuelle d'environ 60% de la capacité de cet incinérateur et de la faible progression des quantités de déchets, nous avons tenu compte, dans l'élaboration des alternatives de disposition des boues de l'usine d'épuration Est, des possibilités d'utiliser l'incinérateur existant de la C.U.Q. et d'y recevoir aussi les boues produites à l'usine Ouest. Notons que l'incinérateur à déchets solides de la C.U.Q. produit actuellement de la vapeur qui est vendue à la Papeterie Reed. Toutefois, avant d'élaborer l'éventail des alternatives possibles, il a fallu s'assurer de la faisabilité d'utiliser l'incinérateur existant de la C.U.Q. pour disposer des boues produites aux deux (2) usines d'épuration.

Faisabilité du procédé

Il existe, actuellement, au moins deux (2) installations en Europe où des boues urbaines sont incinérées conjointement avec des ordures ménagères. Il s'agit de l'usine d'incinération de Wilrijk, en Belgique, et de Lisieux, en France. D'autres installations du même genre sont également en construction en Europe et aux Etats-Unis.

Suite à un voyage d'étude réalisé au cours du mois d'octobre 1982, nous avons recueilli un grand nombre d'informations qui nous permettent d'évaluer les possibilités d'application de ce principe, à l'incinérateur existant de la C.U.Q.

D'une façon générale, il apparaît que l'incinération combinée des boues urbaines avec les ordures ménagères requiert une étape de séchage des boues avant l'incinération et une proportion correspondante de mélange avec les ordures ménagères de l'ordre de 10 à 15% (maximum).

Incinérateur de Wilrijk (Belgique). L'installation de Wilrijk est caractérisée par un incinérateur à grilles mobiles, une tour de refroidissement, un électrofiltre et un sécheur à étages conçu par la firme Seghers Engineering. Les boues de quatre (4) usines d'épuration biologique sont déshydratées à chacun des sites en utilisant, dans trois des quatre cas, des filtres-presses, et dans l'autre (l'usine la plus importante, 75% de la quantité totale de boues), des centrifugeuses. Ensuite, elles sont acheminées par camions au site de l'incinérateur de Wilrijk où elles sont séchées dans un échangeur de chaleur à l'huile thermique, laquelle est chauffée par des fumées provenant de la tour de refroidissement. Après cette étape de séchage, les boues ont une siccité de 80% et dans cet état, elles sont injectées dans le four à ordures ménagères. Le bilan énergétique, entre la quantité de chaleur requise pour sécher les boues et la chaleur produite par leur combustion, est positif, c'est-à-dire, qu'il y a plus d'énergie produite par la combustion que les besoins de séchage en nécessitent. On retrouve, à la figure 1, un schéma de cette installation ainsi que le bilan énergétique.

Incinérateur de Lisieux (France). L'installation de Lisieux, en France, a été conçue par la firme INOR, spécialisée dans l'incinération. Les boues de l'usine de traitement biologique (4,7 t/j de matières sèches) sont acheminées à l'incinérateur à une concentration de 22% où elles sont ensuite asséchées et pulvérisées dans un PULVO-THERM (siccité de 65%) qui utilise des gaz chauds sortant du four, pour réaliser l'échange de chaleur. Le bilan énergétique du procédé est quasi nul (figure 2). Les boues sont ensuite injectées pneumatiquement dans le four à ordures ménagères du type VON ROLL (identique à ceux de l'incinération de la C.U.Q.) pour être brûlées.

Impacts environnementaux. Dans les deux (2) cas, les buées odorantes, produites lors de la phase de séchage des boues, sont retournées dans le four d'incinération où elles sont détruites (température de 800°C). D'après les expériences vécues aux deux (2) endroits (3 à 4 ans à Wilrijk et 1 à 2 ans à Lisieux), la combustion des boues urbaines combinées avec les ordures ménagères ne cause aucun préjudice à l'environnement. De plus, selon des essais réalisés par la Compagnie INOR, les résultats sont identiques, que les boues soient de nature biologique ou physico-chimique.

Application du procédé à l'incinérateur de la C.U.Q. Ainsi, pour l'incinérateur de la C.U.Q., nous pensons à un système de séchage qui utiliserait de la vapeur comme source d'énergie et qui serait situé, pour des raisons pratiques (conduites de vapeur et de retour des buées), sur un terrain localisé au nord du site d'incinération. Les boues pourraient y être acheminées par pompage, pour une déshydratation sur place ou par camions, après déshydratation aux sites des usines Est et Ouest. Les buées odorantes produites par l'assèchement seraient retournées aux tours d'incinération pour y être détruites. En se basant sur des données existantes, en Europe (Wilrijk et Lisieux), et sur certaines valeurs théoriques, le bilan énergétique de l'installation serait:

A) Données de base

- Quantités de boues:

  Est: 24,4 t/j sur une base sèche;

  Ouest: 48 t/j sur une base sèche;

  Evaporation de l'eau: 750 Kcal/kg;

  Pouvoir calorifique inférieur utile des matières sèches: 4 000 Kcal/kg (boues non digérées);

  57% de matières sèches combustibles;

  50% de pertes dans le processus par radiation et autres;

  Incinérateur C.U.Q.:
  - rendement des bouilloires: 70%;
  - 1 220 B.T.U./lb de vapeur;
  - 4,50 $/1 000 lb de vapeur vendues en 1983;
  - 700 t/j d'ordures ménagères actuellement, en moyenne;
  - opération: 7 j/s, 24 h/j.

- Déshydratation mécanique: une siccité de 32% permet d'obtenir un bilan énergétique auto-suffisant. Les machines actuellement sur le marché permettent de couvrir une gamme de siccité comprise entre 25% (centrifugeuses) et 45% (filtres-presses).

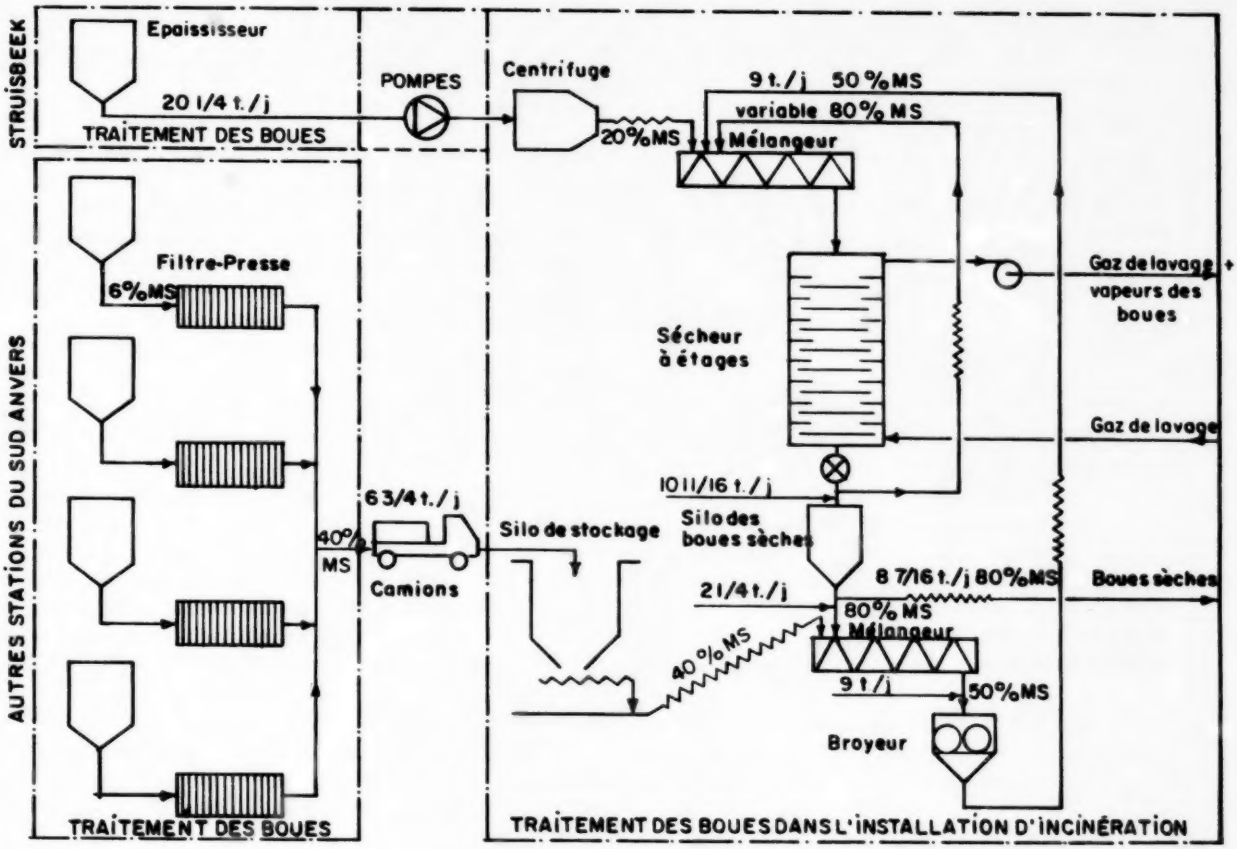
- Assèchement thermique: siccité de 80%.

B) Bilan énergétique - boues du secteur Est

Chaleur d'incinération:	
24,4 t/j x 57% x 4 000 Kcal/kg =	55,6 Gcal./j
moins 20% d'eau dans les boues à 80%	
6 100 kg/j x 750 Kcal/kg =	-4,6 Gcal./j
Chaleur nette d'incinération:	51,0 Gcal./j
moins la chaleur de séchage de 32 à 80% de siccité:	
45,8 t/j d'eau x 750 Kcal/kg + 50% (pertes) =	51,5 Gcal./j
Chaleur nette fournie par les boues:	-0,5 Gcal./j
Pourcentage de boues (siccité de 80%) dans le mélange avec les ordures ménagères:	5%

FIGURE 1

Schéma de Wilrijk



Bilan thermique estimé (assumant 5 710 kcal/kg sur 74% de matière combustible)

- Chaleur générée par l'incinération des boues:

  $8\frac{7}{16}$ t/j x 30% de matières solides x 74% (combustible) x 5 710 kcal/kg = 28,52 Gcal/j

  moins l'eau à évaporer: $8\frac{7}{16}$ t/j x 20% d'eau x 750 kcal/kg = - 1,27 Gcal/j

  Production nette de chaleur à l'incinérateur par les boues = 27,25 Gcal/j

  moins chaleur de séchage:

  eau à l'entrée : 27 t/j x 75% d'eau = 20 250 kg

  eau à la sortie: $8\frac{7}{16}$ t/j x 20% d'eau = 1 687,5 kg

  Quantité d'eau à évaporer = 18 562,5 kg

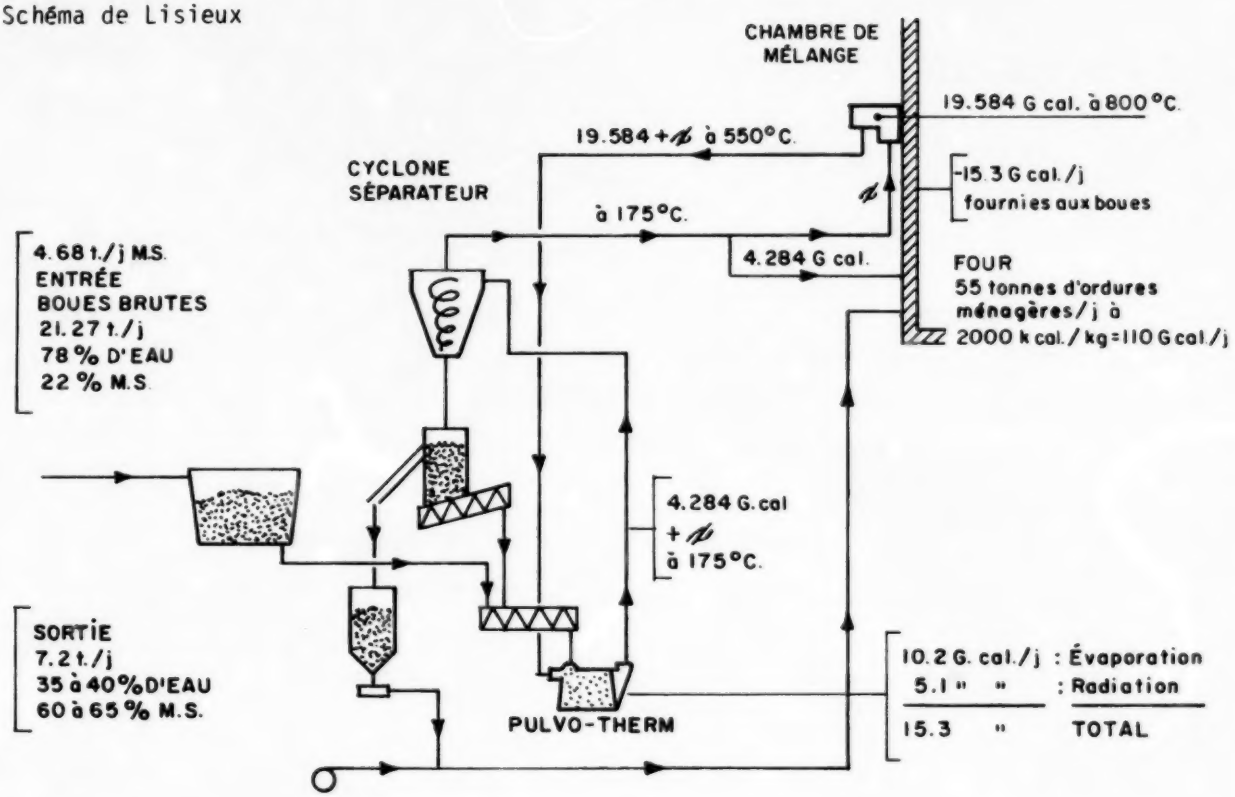
  Chaleur de séchage: 18 512,5 kg d'eau/j à 750 $\frac{\text{kcal}}{\text{kg}}$ = 13,92 Gcal/j

  plus 50% de pertes par radiation = - 20,88 Gcal/j

  Chaleur nette produite par les boues = 6,37 Gcal/j

FIGURE 2

Schéma de Lisieux



Bilan thermique estimé

Si on applique le pouvoir calorifique des boues de Wilrijk à Lisieux:

- Chaleur générée par la combustion des boues:

7 200 kg/j x 60% de matières solides x 75% (combustible) x 5 710 kcal/kg	=	18,25 Gcal/j
moins l'eau à évaporer: 7 200 kg/j x 40% x 725 kcal/kg	=	- 2,088 Gcal/j
Production nette de chaleur à l'incinérateur par les boues	=	16,162 Gcal/j
moins chaleur de séchage: 4,68 t/j (matières solides) x $\frac{78\%}{22\%}$	=	16,59 t/j d'eau
moins 7,2 x 35%	=	- 2,52 t/j d'eau
Quantité d'eau à évaporer:		14,07 t/j d'eau
Chaleur de séchage: 14 070 kg d'eau à 725 kcal/kg = 10,2 Gcal/j plus 50% de radiation	=	- 15,3 Gcal/j
Chaleur nette produite	=	0,862 Gcal/j

C) Bilan énergétique - boues des secteurs Est et Ouest

Chaleur d'incinération =	109,4 Gcal./j
moins 20% d'eau dans les boues à 80% =	9,0 Gcal./j
Chaleur nette d'incinération =	100,4 Gcal./j
moins la chaleur de séchage de 32 à 80% de siccité =	-101,3 Gcal./j
Chaleur nette fournie par les boues =	-0,9 Gcal./j
Pourcentage de boues (siccité à 80%) dans le mélange avec les ordures ménagères =	10%

Ainsi, on constate que le bilan énergétique peut être auto-suffisant pour une siccité des boues de 32%, ce qui est réalisable avec les machines disponibles actuellement, et que la proportion de boues dans le mélange avec les ordures ménagères ne dépasserait pas 10%, si elles sont séchées à 80% avant l'incinération et si elles sont injectées de façon continue dans l'incinérateur.

Compte-tenu du niveau préparatoire de l'étude en cours, des caractéristiques des systèmes en fonction actuellement à Wilrijk et à Lisieux (four VON ROLL comme ceux de l'incinérateur de la C.U.Q.), et des indications fournies par nos calculs exploratoires, il nous apparaît possible d'utiliser l'incinérateur existant de la C.U.Q. pour éliminer les boues urbaines qui seront produites aux deux usines d'épuration (Est et Ouest).

Remarquons, finalement, qu'une évaluation des modifications physiques à apporter à l'incinérateur existant (branchement pour l'alimentation en vapeur pour le retour des buées et pour l'injection des boues sèches dans le four), aurait dépassé le cadre d'une étude de niveau préparatoire.

ALTERNATIVES DE DISPOSITION DES BOUES

Compte-tenu des diverses possibilités de transport (pompage ou camions) et des divers types de procédés applicables, seize (16) alternatives de disposition des boues sont apparues réalisables au cours de l'étude préparatoire. Pour la disposition à l'incinérateur de la C.U.Q., le procédé comprend les étapes principales suivantes:

- emmagasinement des boues fraîches à la station d'épuration (1 journée maximum);
- pompage des boues fraîches du site de l'usine au site de l'incinérateur;
- épaississement des boues au site de l'incinérateur (siccité d'environ 6%);
- déshydratation mécanique sur une base de 7 jours par semaine et 21 heures par jour (siccité de 32%);
- séchage des gâteaux dans un échangeur de chaleur alimenté par de la vapeur (siccité de 80%). Retour des buées odorantes dans les fours d'incinération pour leur destruction;
- injection des boues séchées dans l'incinérateur sur une base continuelle (7 jours par semaine et 21 heures par jour).

Sur la base des informations recueillies lors de nos visites à Wilrijk et Lisieux, et auprès des différents manufacturiers, en appliquant des fonctions de coûts[1] associés aux différents procédés, il fut possible de réaliser une comparaison économique de ces alternatives.

En résumé, la comparaison économique est la suivante:

Description	Construction	Exploitation
A) Secteur Est:		
- Digestion avec enfouissement des boues digérées:	10 400 000 $	1 078 000 $/an
- Incinération dans un incinérateur à boues:	10 000 000 $	1 275 000 $/an
- Incinération à l'incinérateur existant de la C.U.Q. (transport des boues par pompage):	10 100 000 $	787 000 $/an
B) Secteurs Est et Ouest:		
- Incinération à l'incinérateur de la C.U.Q. (transport des boues par pompage):	19 900 000 $	1 430 000 $/an

A la lumière de ces données, on constate que le principal avantage à incinérer les boues conjointement avec les ordures ménagères réside dans une diminution des coûts d'exploitation par rapport aux autres méthodes.

Pour la disposition des boues provenant du secteur Est de la C.U.Q., les écarts de coûts sont les suivants:

- Les frais d'exploitation de la digestion sont de 37% supérieurs à ceux du procédé envisagé;
- Les frais d'exploitation de l'incinérateur conventionnel sont supérieurs de 62%.

On note de plus, que l'addition des boues du secteur Ouest à celles du secteur Est contribue à rentabiliser davantage le procédé par rapport aux autres méthodes. De fait, il n'en coûterait que 12% de plus pour disposer des boues des deux (2) secteurs à l'incinérateur existant, comparativement à l'incinération conventionnelle des seules boues du secteur Est. L'écart atteint 33% lorsque la comparaison est faite avec la digestion.

CONCLUSIONS

Les résultats obtenus au cours de cette première investigation, au sujet de la possibilité de disposer des boues à l'incinérateur à déchets solides de la C.U.Q., sont très encourageants. De fait, le processus peut être autosuffisant pour des boues ayant une siccité de 32% après leur déshydratation mécanique (réalisable avec les machines disponibles actuellement), et les

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1. Hasit, Y., Vesilind, P.A., Journal of Water Pollution Control Federation.

frais d'exploitation annuels sont significativement inférieurs à ceux associés aux autres méthodes.

Ainsi, il apparaît intéressant d'élaborer davantage cette solution au cours des prochaines étapes du projet.

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